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PROCESS EVALUATION OF SELECTED OPTIONS FOR TREATING NEW BRUNSWICK MANGANESE MINERAL DEPOSITS

VOLUME I
RESULTS OF TESTWORK

Project 5211

for

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1.0 SUMMARY

Metallurgical testwork was conducted on the New Brunswick Plymouth manganese ore. Acid leaching, gravity separations, magnetic separations and flotation testwork were performed.

Heap leaching of the coarse (minus 0.95 cm) ore with sulphuric acid and sulphur dioxide at a pH of less than 2.5 yielded a manganese extraction of 71% in approximately 49 days. Optimum pH was 2.0, yielding reagent consumptions of approximately 110 kg/t of sulphuric acid and 90 kg/t sulphur dioxide. The leach solution contained numerous impurities. Future testwork must confirm that conventional purification methods can produce acceptable solution quality.

Oxidizing and sulphation roasting did not yield high manganese extraction from water leaching of the resulting calcines.

Mineral processing was conducted primarily to upgrade the manganese content, and to a lesser extent as a pre-concentration method to reject gangue. Gravity concentration, magnetic separation, and flotation were all successful in concentrating the manganese by a factor of two, however, manganese recoveries ranged widely from 10 to 80%. Further upgading by staged processing (i.e. flotation of gravity concentrate, cleaner flotation) was not successful.

Due to the high reagent cost for leaching, pre-concentration of the ore is essential as a method of gangue rejection. Gravity concentration would be the most economical method, due to its low capital and operating costs relative to flotation. Spiral concentration was the most successful physical method tested. Heavy media testwork may be warranted in view of the success of heavy liquid separation.

1.0 SOMMAIRE

On a effectué des essais métallurgiques sur le minerai de manganèse Plymouth du Nouveau-Brunswick. Il s'agissait d'essais de lixiviation acide, séparation par gravité, séparation magnétique, et flottation.

La lixiviation en tas du minerai grossier (moins 0,95 cm) avec acide sulfurique et gaz sulfureux, à un pH inférieur à 2,5, a permis l'extraction de 71% du manganèse en 49 jours environ. Le pH optimum était de 2,0; dans ces conditions, la consommation de réactifs a été d'environ 110 kg par tonne d'acide sulfurique et 90 kg/t de gaz sulfureux. La solution obtenue par lixiviation contenait de nombreuses impuretés. On devra vérifier par d'autres travaux que les méthodes de purification conventionnelles permettent l'obtention d'une solution de qualité acceptable.

Le grillage avec oxydation et sulfatation n'a pas conduit à une extraction élevée du manganèse par lixiviation à l'eau du minerai calciné ainsi produit.

On a effectué un traitement de valorisation du minerai essentiellement pour améliorer la teneur en manganèse, et en deuxième lieu comme méthode de pré-concentration pour rejeter la gangue. Les opérations de concentration par gravité, séparation magnétique et flottation ont permis de concentrer le manganèse (facteur de 2); cependant, le taux de recouvrement du manganèse a beaucoup varié, de 10 à 80%. On n'a pas pu améliorer la concentration en employant une suite d'opérations (par ex. flottation du concentré séparé par gravité, flottation à un meilleur niveau de propreté).

Du fait du coût élevé des réactifs de lixiviation, la pré-concentration du minerai est l'opération essentielle dans le cadre d'une méthode de rejet de la gangue. La concentration par gravité serait la méthode la plus économique par comparaison à la flottation, du fait du niveau moins élevé de l'investissement nécessaire et des frais d'exploitation. La méthode de séparation physique qui a donné le meilleur résultat était la concentration sur spirales. Compte tenu de la réussite de la séparation en présence d'un liquide dense, il pourrait être judicieux d'effectuer des essais de séparation en milieu dense.

2.0 INTRODUCTION

Upon acceptance of Witteck Proposal No. 136-85 by the Department of Supply and Services, testwork and metallurgical evaluation was started on samples of New Brunswick Plymouth manganese ore.

The objectives of the programme were as follows:

- Production of a manganese concentrate by mineral processing methods; target grade was 35% manganese
- Pyrometallurgical and hydrometallurgical treatment of the concentrate; hydrometallurgical testwork would include leaching and purification to provide a solution suitable for electrowinning
- Optimization testwork to confirm possible flowsheets
- Calculation of order of magnitude capital and operating costs for the selected flowsheet

As the target concentrate grade could not be produced, the focus of the testwork and evaluations was shifted to leaching of the ore. Sulphur dioxide and sulphuric acid in aqueous solution was selected as the lixiviant. The testwork was very encouraging; however, it was curtailed to allow extra monies for economic studies to identify feasible flowsheet options. This would provide a clear direction to investigations under a Phase II contract.

This volume presents the results of the laboratory testwork. The order of magnitude cost estimates are presented in Volume II.

3.0 TEST PROCEDURES

The following procedures were used during the testwork.

3.1 Sample Preparation

Eight barrels of New Brunswick manganese ore were received by Witteck Development on May 9, 1986. The samples were designated as WDI #86-030. The material included five samples from the Plymouth ore body, (P1-P5) and three drums from the New Hartford zone (NH1-NH3).

Three of the Plymouth samples were inspected to determine whether hand sorting was possible. No visual differences were noted, so that neither hand sorting nor ore sorting was pursued.

Plymouth samples 1-4 were sampled for assay and prepared for metallurgical testwork. No sampling of the New Hartford ore was performed.

3.1.1 Plymouth 1 Preparation

Sample P1 was coned four times by shovel and quartered. Two opposite quarters were taken and stage crushed to -6 mesh in a laboratory cone crusher. The ore was riffled and recombined five times, and riffled into 2000 gram and 500 gram portions. Three of the samples were prepared for chemical assay.

On two later occasions, the original ore was reconed, quartered, and two opposite quarters were split. The samples were then stage crushed to 6 mesh, or to a nominal 20 mesh. On each occasion, a 500 gram sample was split out by riffling for analysis.

These samples were used for preliminary grinding, screening, gravity, flotation and magnetic separation testwork.

3.1.2 Plymouth 2, 3 and 4 Preparation

Representative samples from each of the Plymouth 2, 3 and 4 ores were split in the manner described in Section 3.1.1. A sample of each was prepared for assay, and most of the remainder was used in the preparation of Plymouth ore composite P-02.

3.1.3 Preparation of Plymouth Composite P-02

Representative portions of the Plymouth 1, 2, 3 and 4 ore samples were combined and prepared in a manner similar to that described previously. Samples were then crushed in stages using a laboratory cone crusher to either 6 mesh or to 20 mesh dependent on the specific testwork requirements. Two samples were prepared for assay. The remainder was riffled into 500 gram or 2000 gram samples.

These samples were used in the second phase of flotation, gravity and magnetic separation testwork and in extractive metallurgy testwork.

3.1.4 Preparation for Assay

All head and metallurgical samples were prepared for assay as follows. Coarse samples (i.e. 6 mesh or coarser) were reduced either by shatterbox or by pulverizer to approximately 65-80 mesh. The ground sample was riffled and recombined five times, and split in half by riffling. One half was stage ground by shatterbox or pulverizer until 100% passed 150 mesh. The sample was then mixed by rolling, and submitted for assay.

Finer samples (finer than 6 mesh) were similarly treated with the exception of the initial size reduction.

3.2 Chemical Analyses

Analyses were performed in the following manner:

Liquids: Direct analysis by Atomic Absorption Spectrograph (AA) or by Direct Current Plasma (DCP)

Solids: Acid Dissolution followed by AA or DCP analysis

In addition, a 52 element quantitative analysis was performed on samples of Plymouth P1 ore and flotation products. This analysis involved neutron activation, x-ray fluorescence, and acid dissolution/instrumental analysis techniques.

3.3 Screen Analyses

Representative samples were shaken by RoTap for forty-five minutes through sieves as fine as 200 mesh.

Determination of size distributions for -200 mesh material was performed by wet sieving through a 400 mesh sieve. The +400 mesh material was then dried, and shaken by RoTap for thirty minutes through 200, 270, 325 and 400 mesh sieves. An overall particle size distribution was then calculated.

In some cases, each size fraction was prepared and submitted for analysis.

3.4 Bond Index Determination

Four of the 2000 gram samples of the P-02 composite were utilized as feed for the Bond Index test work. The mill used was a Bico brand ball mill with a 20 kg charge of 1.75 cm to 3.75 cm steel balls. The screen used was 150 mesh (106 microns) and the mill operation brought to an equilibrium circulating load of 250 percent.

The Work Index is a comminution factor which expresses the resistance of a material to grinding. It is defined as the number of kilowatt hours required to reduce one tonne of material from theoretically infinite size to 80% passing 100 microns (or about 67% passing 200 mesh).

The equation defining the Work Index is:

Bond Formula

$$Wi = 49.1 \,/\, (P1)^{0.23} \times (Gbp)^{0.82} \times \left(\frac{10}{\sqrt{P}} - \frac{10}{\sqrt{F}}\right)$$

Where: Wi = Work Index (kwh/tonne)

P1 = Screen size test in microns

Gbp = Net grams of undersize produced per revolution of test mill

P = Size in microns which 80 percent of test product passes

F = Size in microns which 80 percent of test feed passes

Using this Work Index equation, Wi should conform with the motor output power of an average overflow ball mill of 2.44 m internal diameter grinding wet in closed circuit.

The Work Input is a measure of the power required (in kilowatt hours per tonne of ore) for the screen size tested. The defining equation is:

$$W = 10\,Wi \left(\sqrt{\frac{1}{P}} - \sqrt{\frac{1}{F}} \right)$$

The Work Index is a function of the diameter of the mill utilized. The following correction factor should be applied in order to provide for mill diameter.

$$\text{Factor} = \left(\frac{2.44}{D} \right)^{0.20}$$

Where D = diameter in meters of the mill to be considered.

3.5 Gravity Testwork

Various methods of gravity treatment of Plymouth ore samples were tested. They included heavy liquid separation, vanning, jigging, tabling and spiralling. Gravity test feeds were prepared as follows.

1. The 6 mesh P1 ore was ground in a 35 cm x 20 cm diameter mill with steel rods or steel balls, or a 20 cm x 20 cm diameter steel ball mill for vanning, tabling and spiralling.

2. The +65 mesh fraction of P1 ore, ground as described above, was used for jig testwork.

3. Screen fractions of P1 ore, ground as described above, were used in heavy liquid separation testwork.

4. The six mesh P1 or P-02 ore composite was stage crushed to a nominal -20 mesh (i.e. 97-99% -20 mesh) for spiral testwork.

5. The +65 mesh fraction of the -20 mesh P-02 ore composite was used for spiral testwork.

3.5.1 Heavy Liquid Separation

Heavy liquid separations were performed with tetrabromoethane at a specific gravity of 2.96 g/cc. The ground P1 ore was screened and the +14, -14 +28, -28 +48, -48 +65, -65 +100 and -100 mesh fractions were tested. A sample of twenty grams of each fraction was placed in the liquid in a 250 mL glass cone, stirred, and allowed to separate. The sink fraction was removed from the apex of the cone and the sink and float fractions were filtered, washed in acetone and water, dried and prepared for assay.

3.5.2 Vanning Testwork

Samples of the ground P1 ore were vanned with a mechanical vanner, and with a 25 cm diameter vanning plaque. The concentrate of each test was filtered, dried, and prepared for assay.

3.5.3 Jig Testwork

Recovery of gold by jigging was performed with a Denver laboratory size jig. Dry ore was fed to the jig using a vibratory feeder. The +65 mesh fraction of the ground P1 ore was used as feed.

Stroke length and water flow rate were adjusted to provide optimum operating conditions.

The jig products were filtered, dried and prepared for assay.

3.5.4 Table Testwork

Two table tests were performed on ground P1 ore. In each test, ten kilograms of ore was slurried to approximately 25% solids in a 27 cm diameter conical bottom slurry tank and fed by gravity to a 1.9 m x 1.4 m Deister shaking table to concentrate the manganese minerals. Pitch, stroke length, and frequency were adjusted until suitable products were obtained. Products were collected in wooden launders and flowed into buckets. Each product was later dewatered, filtered, dried, weighed and prepared for analysis.

3.5.5 Spiral Testwork

The ground or crushed P1 or P-02 ore was slurried in an 80 litre capacity Denver slurry tank at 20-35% solids, and pumped by an Allis-Chalmers 5 cm x 5 cm SRL pump to the spiral feed port. Flow rate was controlled by means of a return line to the tank.

The spiral used was a MK7 Reichert spiral. It was thought that a Reichert spiral would provide a better separation than a Humphrey's spiral for this particular ore.

The slurry was split into a concentrate, middlings, and tailings by adjusting the splitters. The product flowed back into the slurry tank and when conditions were as desired, samples were split.

3.6 Magnetic Separation

Low and high intensity magnetic separation of the ground P1 and P-02 ores were tested.

3.6.1 Low Intensity

A sample of the ground P-1 ore was passed through a Sala 200 mm diameter x 100 mm drum, low intensity wet magnetic separator. The ore was fed by beaker as a 25% solids slurry. As very little magnetic material was recovered, samples were not taken for assay.

3.6.2 High Intensity

Two sets of tests were performed with wet high intensity magnetic separators on P-02 ore composite. Test conditions are outlined below.

WHIMS I
- Box mag high intensity magnetic separator
- Coarse matrix (approx. 6-8 mm)
- Field strength 4000, 11000 and 16000 gauss
- Feed passed at 4000 gauss, non magnetics repassed at 11000 gauss, non magnetics passed again at 16000 gauss
- Feed 78% -280 mesh at 33 mL/sec

WHIMS II
- Sala high intensity magnetic separator
- Fine matrix (approx. 1 mm)
- Field strength 4200-6800 gauss
- Separate screen fractions, +200, -200 + 400, and -400 mesh were treated with appropriate matrices
- Two separations were made on the coarse sample, three on the medium size, and four on the fine material.

3.7 Flotation

Flotation tests were performed in a Denver D-12 flotation machine. A 500 g cell with a small, closed impellor was generally used. In some of the later tests, cleaner flotation was performed in a 250 gram cell. Conditioning was performed at 33-40% solids. Flotation was generally conducted at 20-30% solids.

Flotation was either direct flotation of manganese minerals, or reverse flotation of gangue. Most tests were performed on Plymouth P1 and P-02 ore samples. Two tests were performed on the -65 mesh fraction of the crushed P-02 ore. Three tests were conducted on spiral products.

3.7.1 Grinding

Grinding for flotation was performed at 65% solids in a 20 cm x 20 cm steel ball mill containing 10 kg of 1.9 - 3.8 cm steel balls. Five hundred grams of ore were ground in each batch. In some of the tests, the ore was ground, screened through a 65 mesh sieve, and the oversize was reground in the same mill. This stage grinding was performed to reduce sliming.

Upon completion of grinding, the slurry was washed into the flotation cell, or into a desliming vessel.

3.7.2 Desliming

In some tests, the ground slurry was deslimed prior to flotation. The slurry was agitated in a 20 L bucket at 5% solids or in a 4 litre beaker at 12% solids. A dispersant was added, and after twenty minutes the agitation was stopped.

The solids were allowed to settle, and the slimes decanted by siphoning. The procedure was repeated without additional dispersant, until most of the slimes were removed.

Stokes' Law was used to calculate the required time to produce slimes of a certain size. However, the obtained products were generally much heavier or lighter than desired.

3.7.3 Reverse Flotation

Reverse flotation was performed as described previously, with the following special conditions.

1. Sodium metasilicate was not used as a dispersant, as it is a silica depressant. Instead, sodium hexametaphosphate (HMP) was the dispersant.

2. Dextrin was used in most tests as a manganese depressant.

3. Flotation pH was adjusted to 10-11 with caustic soda.

4. In some tests, the reagents in the flotation tail were oxidized, and subsequent direct flotation was conducted.

3.7.4 Direct Manganese Flotation

Direct flotation was performed as described previously, with the following special conditions.

1. Sodium metasilicate was used in most tests as a silica depressant, and in desliming as a dispersant.

2. Flotation pH was adjusted with hydrochloric acid or caustic soda in some tests. Most tests were, however, conducted at ambient pH. Addition of sodium metasilicate raised the pH to 9-10. However, the pH subsequently decreased to 7.5 - 8.0 during flotation.

3. In one test, the reagents in the flotation concentrate were oxidized with hypochlorite, and reverse flotation was conducted.

3.7.5 Reagents

The slurry in the flotation cell was adjusted to 33-40% solids. The pH was modified, if desired, and reagents added. In some tests, existing reagents were oxidized with hypochlorite prior to the next stage of flotation. In two tests, the ore was sulphidized with sulphur dioxide prior to flotation.

The following lists reagents used.

Dispersants

Sodium hexametaphosphate (HMP)	Dissolved Solid
Sodium metasilicate	Dissolved Solid

pH Modifiers

HCl	1 N or 5 N
NaOH	1 N
Na_2CO_3	Solid

Frothers

AeroFroth 65	52 gpl
Pine Oil	Undiluted Liquid
Fuel Oil	Undiluted Liquid

Manganese Collectors

Oleic Acid	100 gpl in methanol

Gangue Collectors

Aero Float 845 (petroleum sulphonate for calcite)	10 gpl
Aero Float 899 (petroleum sulphonate for Fe oxides)	10 gpl
Armac C (amine for silica and silicate)	10 gpl

Depressants

Sodium metasilicate (for silica)	Dissolved solid
Dextrin (for manganese minerals)	Dissolved solid
Quebracho (for manganese minerals)	Dissolved solid

Oxidizers

Calcium hypochlorite	Dissolved solid

Sulphidizers

Sulphur dioxide	Gas

3.8 Mineralogy

Samples of flotation and gravity test products were selected for petrographic examination. Polished thin sections were prepared for optical microscopy in order to determine the identity of manganese and gangue minerals, their associations, and degree of liberation in the concentrates and tails.

The sections were examined with an optical microscope in transmitted and reflected light, using magnifications of 50, 100, 200, 500 and 1250 x, the latter with oil immersion.

3.9 Extractive Metallurgy

Twelve tests were conducted to investigate methods of manganese dissolution. Three preliminary tests were conducted on flotation products. One heap leach and eight bottle leach tests were performed on Plymouth ore samples.

3.9.1 Flotation Products

Three exploratory tests were conducted on a sample of the combined flotation products of Tests MNF 14-19. The tests included:

- Sulphatizing roast with Na_2SO_3, followed by water leaching.
- Oxidizing roast with H_2SO_4 and NaCl, followed by leaching in water.
- Reducing leach with SO_2 and H_2SO_4

The leaches were conducted at low pulp densities (less than 1% solids) and with very high reagent levels to ensure maximum Mn extractability. Test conditions are outlined in Appendix VII.

3.9.2 Column Heap Leach of Ore

A column heap leach was conducted on the Plymouth P-02 ore. The ore was first crushed to -0.95 cm, agglomerated with sulphuric acid, cured, loaded into a 5 cm diameter plastic column, and leached with a circulated H_2SO_4/SO_2 solution. Test conditions are presented in Appendix VII.

3.9.3 Bottle Leaches of Ore

A total of 8 bottle leaches were conducted on the -0.95 cm Plymouth P-02 ore. The ore was put into a 2 litre Winchester with a solution of H_2SO_4 and sodium metabisulphite ($Na_2S_2O_5$), and rolled at approximately 28 rpm for 5-14 days. Test conditions are detailed in Appendix VII.

4.0 RESULTS AND DISCUSSION

The following results were obtained from the testwork.

4.1 Literature Search

A literature search was conducted to obtain data which would help to define test parameters. Papers obtained are detailed in Appendix I.

4.2 Head Assays

The head assays for the Plymouth ore samples are summarized in Tables 1 and 2. The P1 zone manganese assays ranged from 6.7-7.6%.

The P-02 composite was made from 11 kg of P1 ore, 142 kg of P2, 66 kg of P3 and 148 kg of P4. The calculated head assays of the composite were 7.60% Mn and 10.02 of Fe. The assayed values were 7.57% Mn and 9.46% Fe.

The P1 ore contained 38.7% SiO_2 and 14.8% Al_2O_3. Other elements of interest included TiO_2 (0.49%), Ta (less than 0.5 ppm), Nb (20 ppm), Y (20 ppm), Zr (70 ppm) and the rare earth elements Yb (2.23 ppm), Eu (1.26 ppm), Tb (0.8 ppm), Lu (0.34 ppm) and Sm (5.4 ppm).

Table 1

MANGANESE AND IRON ASSAYS OF PLYMOUTH ORE SAMPLES

Ore Sample	Designation	Sample No.	Mn (%)				Fe (%)		
			1	2	3	Avg.	1	2	Avg.
P1	P1-I	A	6.27	6.36	6.29	6.31	6.56	7.24	6.90
		B	6.91	6.56	6.15	6.54	6.55	7.00	6.78
		C	7.37	6.43	6.48	6.76	6.81	6.88	6.84
		Avg.				6.54			6.84
P1	P1-II	A	6.76	-	-	6.76	7.03	-	7.04
P1	P1-III	A	7.66	-	-	7.66	8.95	-	8.95
P2	P2	A	7.48	-	-	7.48	7.11	-	7.11
P3	P3	A	6.21	-	-	6.21	8.91	-	8.91
P4	P4	A	8.41	-	-	8.41	13.55	-	13.55
P1-P4	P-02	A	7.62	-	-	7.62	9.50	-	9.50
		B	7.52	-	-	7.52	9.41	-	9.41
		Avg.	7.57			7.57	9.46		9.46
P1-P4 (+65 mesh)	P-02 (+65 mesh)	A	7.62	-	-	7.62			
P1-P4 (-65 mesh)	P-02 (-65 mesh)	A	6.65	-	-	7.57			

P1-I	Used for grinding, vanning, heavy liquid and jigging testwork
P1-II	Used for table testwork and spiral tests 1-5
P1-III	Used for spiral tests 6 & 7 and flotation tests MNF 1-19

Composite P-02 - Used for flotation tests 22-33

Composite P-02 +65 mesh - Used for spiral tests 8-14

Composite P-02 -65 mesh - Used for flotation tests 34 and 35

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Table 2

SEMI-QUANTITATIVE ANALYSIS OF PLYMOUTH ORE *

Element	Weight %	Element	ppm	Element	ppm
SiO_2	38.7	Ba	1830	Th	7.4
Fe_2O_3	14.8	Sr	260	Be	6.0
Al_2O_3	11.0	Cr	120	Sm	5.4
MnO	9.69	Rb	100	Cs	4.3
CaO	6.65	Ni	78	Pb	4.0
K_2O	2.59				
MgO	2.50	Co	77	Hf	3.1
Na_2O	0.91	As	77	W	3.0
		Zn	76		
		Zr	70	Yb	2.23
P_2O_5	0.55	Ce	70	U	2.2
TiO_2	0.49	Cu	46	Mo	2.0
L.O.I.	11.3	Li	46	Br	0.8
		B	40	Se	1.4
Total	99.2	La	31.1	Eu	1.26
		V	28	Bi	0.9
		Nd	25	Tb	0.8
		Nb	20	Lu	0.34
		Y	20	Ta	LT 0.5
		Sc	14.7	Ag	LT 0.5
		Ge	LT 10.0	Cd	LT 0.2
		Sb	7.8	Au	0.093

* Pl ore

LT - less than

4.3 Ore Grindability

4.3.1 Bond Index

The Bond Index of the -6 mesh P-02 ore was 15.93 kwh/tonne. The work input was 14.46 kwh/tonne for grinding to 80% -80 microns. The ore was therefore fairly hard despite the fact that a fraction was very easily ground to slimes.

Detailed test data are shown in Appendix II.

4.3.2 Size and Metal Distributions

The particle size distributions of the as received Plymouth P1 ore, and of the -6 mesh P1 ore are summarized in Tables 3 and 4. The +3 mesh and -6 mesh fractions comprised 56.1% and 20.1% respectively of the as received ore.

Metal distributions for ground P1 ore are shown in Table 5. The manganese assay for each fraction coarser than 400 mesh was greater than 6%. The -400 mesh fraction assayed less than 5% Mn. While the finer material was slightly lower grade, no preconcentration of the ore by screening or cycloning was feasible.

It should be noted that the relatively brief grinding of the ore produced a product which was 30% -400 mesh and 41% -200 mesh. A component of the ore was very soft and ground easily. The over production of slimes by grinding was therefore of great concern.

All of the particle size data from all of the sieve analysis testwork are shown in Appendix III.

Table 3

PARTICLE SIZE DISTRIBUTION OF AS RECEIVED PLYMOUTH P1 ORE

Size (mesh)	Weight %	Cummulative % Ret.	Cummulative % Pass.
3	56.1	56.1	43.9
4	14.3	70.4	29.6
6	9.5	79.9	20.1
8	6.0	85.9	14.1
10	3.5	89.4	10.6
35	5.7	95.1	4.9
65	2.0	97.1	2.9
100	0.4	97.5	2.5
150	0.4	97.9	2.1
200	0.4	98.3	1.7
Pan	1.7	100.0	0.0
Total	100.0		

Table 4

PARTICLE SIZE DISTRIBUTION
MINUS 6 MESH PLYMOUTH P1 ORE

Size (mesh)	Weight (%)	Cummulative %	
		Ret.	Pass
14	33.4	33.4	66.6
28	20.8	54.2	45.8
48	13.4	67.6	32.4
65	5.0	72.6	7.4
100	3.9	76.5	23.5
Pan	23.5	100.0	0.0
Total	100.0		

Table 5

PARTICLE SIZE AND METAL DISTRIBUTIONS OF GROUND P1 ORE

Size	Wt	Assay (%)		Units (g)		% Distribution		
Mesh	(g)	Mn	Fe	Mn	Fe	Wt	Mn	Fe
+65	150.8	7.15	8.24	10.78	12.43	31.32	36.39	38.24
-65 +100	37.8	6.01	6.60	2.27	2.49	7.85	7.66	7.66
-100 +150	42.4	6.32	7.08	2.68	3.00	8.80	9.04	9.23
-150 +200	45.7	6.57	6.54	3.00	2.99	9.49	10.12	9.20
-200 +325	44.4	6.50	6.05	2.89	2.69	9.22	9.75	8.27
-325 +400	14.2	6.54	5.88	0.93	0.83	2.94	3.14	2.55
-400	146.2	4.84	5.53	7.08	8.08	30.38	23.90	24.85
TOTAL	481.5	(6.15)	(6.75)	29.63	32.51	100.00	100.00	100.00

Feed was -6 mesh, ground in 500 gram ball mill at 65% solids for 5 minutes.

4.4 Gravity Testwork

The gravity testwork consisted of heavy liquid separation, vanning, jigging, tabling, and spiral separation. A summary of the gravity results is shown in Table 6. Detailed mass balance data are shown in Appendix IV.

Maximum concentrate grade by any method was 17.5% Mn by the heavy liquid separation. The maximum grade obtained by any of the process equipment was 14.57% Mn in the spiral testwork.

4.4.1 Heavy Liquid Separation

The results of the heavy liquid separation testwork are summarized in Tables 6 and 7.

The -6 mesh ore was ground in the 35 cm x 20 cm diameter rod mill for two minutes at 80% solids. The product was 45.7% -28 mesh and 23.5% -100 mesh.

The sink of each size fraction coarser than 100 mesh was 16.0 -19.5% Mn grade with a recovery of greater than 80% in 27-37% of the weight. Very little of the -100 mesh material was recovered in the sink fraction. Iron followed a similar trend. This confirmed that gravity concentration of manganese from the ore was theoretically feasible.

4.4.2 Vanning Testwork

The ore for the vanning testwork was ground in the 20 cm x 20 cm diameter ball mill at 65% solids for five minutes. The ore was 68.6% -65 mesh and 42.5% -200 mesh.

The results of the vanning tests are shown in Table 6. Some concentration was obtained with the vanning plaque.

Table 6

SUMMARY OF PRELIMINARY GRAVITY TESTWORK *

					Product				
			Tail Assay		Assay		% Distribution		
Test	Feed	Recovery Basis	% Mn	% Fe	% Mn	% Fe	Wt.	Mn	Fe
Heavy Liquid	Ground Ore	Sink	2.65	3.86	17.51	14.20	25.65	69.66	55.93
Van Test V1	Ground Ore	Conc.	-	-	9.62	-	-	-	-
Van Test V2	Ground Ore	Conc.	-	-	6.78	-	-	-	-
Jig Test J1	Ground +65 mesh fraction	Conc. + Bed	9.35	8.21	10.87	13.43	17.31	19.57	25.50
Jig Test J2	Ground +65 mesh fraction	Conc. + Bed	5.98	5.86	11.02	12.03	28.37	42.19	44.85
Table Test T1	Ground Ore	Conc. only Conc.+Middling	5.90 4.05	6.23 5.22	12.08 9.31	9.25 7.03	21.37 60.45	35.72 77.84	28.25 69.97
Table Test T2	Ground Ore	Conc. only Conc.+Middling	5.22 4.41	5.79 5.54	8.24 7.96	8.18 7.05	33.59 69.50	44.41 78.42	41.92 74.24
Spiral Test MSP-07	Crushed Ore	Conc. only Conc.+Middling	7.57 5.04	- -	14.00 14.57	- -	1.79 27.48	3.27 52.28	- -

* Displayed results for spiral testwork are best obtained results; all other results are for all conducted tests.

Table 7

SUMMARY OF HEAVY LIQUID METAL RECOVERIES ON GROUND ORE P1

Size Mesh	Sink Assay (%) Mn	Sink Assay (%) Fe	Sink % Recovery Wt	Sink % Recovery Mn	Sink % Recovery Fe
+14	16.01	13.38	36.68	86.97	73.74
-14 +28	19.49	15.08	31.79	84.55	69.63
-28 +48	19.02	15.43	30.10	85.05	67.84
-48 +65	17.06	14.68	28.28	80.59	63.65
-65 +100	19.38	15.00	27.67	85.94	70.51
-100	10.03	10.90	1.13	2.11	2.07
TOTAL	(17.51)	(14.20)	25.65	69.66	55.93

Feed was -6 mesh ore ground in 2000 gram rod mill for 2 minutes at 80% solids.

4.4.3 Jig Testwork

Ore for the jigging tests was ground as for the vanning tests. Jig feed was the +65 mesh fraction. Test results are summarized in Table 6. Maximum product grade was 12.0% Mn. Mn recoveries from the jig feed were 35.7-42.1% in 28-45% of the weight.

4.4.4 Table Testwork

Table test feed was ground as for the heavy liquid separation in the 35x20 cm rod mill for two minutes at 80% soldids. The resultant feed was 45.7% -28 mesh and 23.5% -100 mesh. Test results are summarized in Table 6.

The best test yielded a concentrate grade of 12.0% Mn. Recovery was 35.7% Mn in 21.3% of the weight. Recovery in the combined concentrate and middling was 77.8% in 60.4% of the weight. The product grade was, however, reduced to 9.3% Mn.

In summary, jigging and tabling were unsuccessful in providing upgrading similar to that achieved in the heavy liquid testwork. Further gravity testwork was conducted with a Reichert spiral. The spiral was expected to be more suitable than the table for treatment of this ore where a relatively small differential in specific gravities of managanese and gangue minerals exists.

4.4.5 Spiral Testwork

Four individually prepared feed samples were used in the spiral testwork, as summarized in Table 8. Test results are shown in Table 9.

Table 8

FEEDS USED IN SPIRAL TESTWORK

Spiral Test #	Ore Type	Ore Mesh	Details	Cum. Wt. % Passing
MSP 01-05	P1	6	Ground 5 min. at 80% solids in 35x20 cm diameter rod mill; +28 mesh reground in same mill to 100% -28 mesh	100% -28 mesh 73% -35 mesh 30.2% 100 mesh
MSP 06-07	P1	20*	Crushed ore	80.9% -28 mesh 65.2% -35 mesh 18.9% -100 mesh
MSP 08-12	P-02	20	+65 mesh fraction of Crushed ore (finer than for MSP 06-07)	75.4% -28 mesh 59.6% -32 mesh 19.7% -48 mesh
MSP 13-14	P-02	20	As for MSP 08-12 with approx 25% of -65 mesh fraction included.	78.8% -28 mesh 65.2% -32 mesh 9.1% -100 mesh

* Nominal

A high slurry flowrate and relatively high solids loading on the spiral were expected to provide optimum results for this ore. The best test was test MSP 07, which produced a combined concentrate and middling of greater than 14% grade and 52.2% recovery in 27.4% of the weight. It is interesting to note that the corresponding concentrate, which represented only 1.7% of the weight, also graded 14.0% Mn. Therefore, no concentration differential existed across this width of the spiral.

The initial seven tests were conducted on crushed and ground P1 ore. The subsequent tests were performed on the +65 fraction of crushed -20 mesh Plymouth composite P-02 ore composite, as it was suspected that the fine material on the spiral might result in a poor overall separation. Spiralling in these final seven tests was not as successful as that of Test 07. The best results of Tests MSP 08-14 were obtained in Test 13. Manganese grade was 12.17%. Recovery was 41.0% in 25-37% of the weight.

The concentrate, and combined concentrate and middling of Test MSP 14, were used in subsequent flotation testwork.

The testwork suggested that gravity treatment of the ore by spiralling could result in a doubling of the ore grade with reasonably high recoveries. Further upgrading was presumably not possible due to the mineral associations and degree of liberation. Effective liberation would not be achieved without size reduction which would be excessively fine for gravity techniques. The small differential in specific gravities between managanese and gangue minerals was also a contributing factor.

Spiral treatment of the P1 ore resulted in higher recoveries than for the P-02 composite. The two feeds were not equivalent as the -65 mesh fraction of the P-02 ore was screened and was not fed to the spiral.

Table 9

SUMMARY OF SPIRAL TESTWORK ON PLYMOUTH ORE

Test No.	Purpose	Feed				Concentrate			Concentrate + Middling		
		Type	%Solids	L/min*	kg/min*	Wt%	%Mn	% Mn Dist	Wt %	% Mn	% Mn Dis.
MSP-01	Lower solids content	1	25-30	35	12	10.66	13.20	20.83	59.02	8.59	75.12
02	Higher solids, flow	1	30-35	86	25	7.33	13.16	15.00	43.37	10.21	66.16
03	Lower flow than 02	1	30-35	60	17	10.47	12.95	23.34	48.84	7.39	62.12
04	Decrease Middlings wt.	1	25-30	39	13	9.92	15.49	22.88	27.48	10.70	43.76
05	Decr. middl. wt. of 04	1	25-30	31	13	9.92	12.28	22.34	21.37	8.56	33.55
06	Higher flow, incr.conc.	2	25-30	68	21	6.80	13.60	12.07	42.72	9.56	53.29
07	Low wt. conc.	2	25-30	60	21	1.79	14.00	3.27	27.48	14.57	52.28
08	As MSP 07	3	18-20	NA	NA	13.91	10.04	18.33	54.33	7.78	55.46
09	Low pulp density	3	25-28	51	NA	10.26	7.54	10.15	30.77	9.73	39.28
10	Higher pulp density	3	25-28	60	NA	9.39	12.63	15.57	40.88	8.93	47.93
11	As MSP 10, lower wts.	3	25-28	60	NA	1.66	13.86	3.01	22.65	12.05	35.83
12	As MSP 07	3	25-28	73	NA	1.81	13.71	3.26	14.93	10.23	20.05
13	Finer feed	4	25-28	67	NA	9.45	13.17	16.55	25.37	12.17	41.05
14**	As MSP 13	4	25-28	67	21	14.16	11.17	21.38	25.72	10.54	36.63

Feeds: 1 - ground P1 ore 2 - crushed P1 ore 3 - crushed P-02 composite, + 65 mesh
4 - crushed P -02 composite with 25% of -65 mesh included

* Approximate, accurate measurements were difficult to obtain with the equipment used.
** Concentrate, and concentrate-middlings composite used in flotation testwork.
0853A

4.4.5.1 Mineralogy of Spiral Products

Similar particle types were observed in each of the concentrate, middlings and tailings products from the spiral testwork. However, the ratios of specific particle types exhibited wide variations between products. Photomicrographs of the spiral products are shown in Appendix VI.

The spiral concentrate contained numerous high density fragments including massive, coarse-grained, liberated manganese oxides, chiefly manganite, pyrolusite, braunite and psilomelane and intergrowths of Mn-Fe-Ti oxides with gangue minerals. The latter particle type typically contained abundant oxides with fine-grained silicates and probably manganese carbonates. Accessory Fe oxide and gangue fragments comprised hematitic slates, quartzofeldspathic chlorite-carbonate particles and lesser coarse carbonates as free and intergrowth type non-liberated grains.

The middlings products exhibited an increase in gangue mineral content, primarily due to non-liberation of manganese oxides. Intergrowth-type particles contained less than 40-50% manganiferous phases. Hematitic slate, liberated carbonates and silicates (chlorite, quartz-feldspar) exhibited an increased abundance as compared to the concentrate.

The tailings product contained minor to trace quantities of coarse liberated manganese oxides while the majority of the manganese was restricted to intergrowths of carbonates-silicates and oxides i.e, non-liberation of the manganese-bearing grains. The manganese formed as low oxide content intergrowths and also as low density and/or Mn-poor minerals such as chlorite.

In summary, the spiral products exhibited a progressive decrease from concentrate to tailings in coarsely liberated Mn-oxides and in oxide to silicate-carbonate ratios of non-liberated intergrowth particles. A segregation of individual manganese species by average specific gravity, though limited by average degrees of liberation, was observed i.e. oxide-dominant concentrate, carbonate-dominant middlings and silicate-rich tailings.

Gravity treatment was therefore successful in concentrating particles comprised of Mn-rich minerals, primarily oxides. The relatively poor grade (less than 14% Mn) was confirmed by the apparent density segregation and poor degree of liberation of Mn-bearing species.

4.5 Magnetic Separation

4.5.1 Low Intensity

One test was conducted with a low intensity magnetic separator. The feed was ground for two minutes in the 35 cm x 20 cm diameter rod mill at 80% solids, producing a 45.7% -28 mesh and 23.5% -100 mesh product. As very little magnetic material was separated, no assays were performed.

4.5.2 High Intensity

The results of the high intensity magnetic separation are summarized in Table 10. Detailed mass balances are shown in Appendix V.

Table 10

HIGH INTENSITY MAGNETIC SEPARATION OF PLYMOUTH ORE COMPOSITE

Test	Field Strength (gauss)	Nonmagnetics Assay, %		Magnetics (Cumulative Results)				
				Assays, %		% Distribution		
		Mn	Fe	Mn	Fe	Wt	Mn	Fe
WHIMS								
I	LT 4000*	6.64	7.86	11.23	11.70	18.47	27.70	25.23
	LT 11000	5.43	6.93	11.03	11.39	36.75	54.13	48.85
	LT 1600	4.16	5.86	10.49	11.01	52.61	73.68	67.59
II	4200-6800	2.92	-	13.83	-	35.00	71.87	-

* 4, 10 and 20 amps = 4,000, 11,000 and 16,000 gauss respectively

LT - less than

The ore was ground in a 35 cm x 20 cm diameter ball mill for 30 minutes at 65% solids. The resulting ore was 96.8% 200 mesh and 77.7% -270 mesh.

In the initial testwork (Whims I) with a coarse matrix, magnetic product grades were less than 12%. Recoveries were similar to those achieved in table and spiral testwork.

Overall results in the Whims II test series, with a fine matrix were only slightly better. Maximum product grade in one test on the +200 mesh fraction was 14.8% Mn. Maximum product grade was obtained with the lowest field strength (4.2 kg) and the highest flow velocity (144 mm/sec). These conditions produced a product grading 13.83% Mn, with a manganese recovery of 71.8% in 35.0% of the weight. Decrease in flowrate and increase in field strength resulted in an increase in recovery, and decrease in grade.

These results were not very encouraging. Variation of flowrate, and cleaning may result in higher product grade. However, further testwork was not planned.

4.6 Flotation

Thirty-six flotation tests were performed on the Plymouth ore and spiral products. Flotation of the ore was conducted to develop possible reagent regimes and to optimize parameters prior to flotation of gravity concentrates.

4.6.1 Direct Flotation of Manganese Minerals from the Ore

Thirteen tests were conducted with oleic acid to directly float manganese minerals .

Feed for Tests MNF 1-16 was P1 ore crushed to a nominal 20 mesh (80.9% -28 mesh, 18.9% -100 mesh). Feed to MNF 30 and 33 was P-02 ore crushed to -20 mesh (85.1% -28 mesh, 39.4% -100 mesh).

The results are summarized in Table 11. Maximum concentrate grade was obtained in Test MNF 15. Concentrates floated from 0-4.5 minutes, 4.5-7.5 minutes, and 7.5-10.5 minutes yielded grades of 6.50, 16.89 and 10.42% Mn respectively. Overall grade was 9.79%. Manganese recoveries in Tests MNF 14-16 and 33 ranged from 42.7-57.9%. Increased recovery generally resulted in lower overall grade.

4.6.1.1 Effect of Grind

The grind/particle size was varied in Tests MNF 01, 02, 06 and 30 from 16% to 66% -200 mesh. The data did not conclusively indicate that finer grinding improved Mn grade or recovery.

4.6.1.2 Effect of pH

The pH was varied in Tests MNF 09, 04, 05 and 08. These tests were conducted without sodium metasilicate or desliming. Acidic pH resulted in very low recoveries. A very basic pH resulted in a slightly increased recovery over that achieved at the ambient pH. Subsequent tests were conducted at the ambient pH.

Table 11

SUMMARY OF DIRECT ORE FLOTATION TESTS (2)

Test #	Purpose	Grinding		Wt %	Flot'n Time		Reagents (kg/t)				(1) Tail	Calc Head	Concentrate		
		min	%-200 mesh	Deslimed	(min)	Oleic	NaSilic.	NaOH	HCl	Frother	%Mn	%Mn	%Wt	%Mn	%Mn Dist
MNF-01	Mn Flot'n	4	16.0	-	11.0	2.50	1.50	-	-	-	6.08	7.13	28.41	9.78	38.97
02	Finer Grind	5	41.2	-	11.0	2.50	1.50	-	-	-	6.51	7.37	30.05	9.36	38.18
06	Stage Ground	4/3	35.3	-	11.0	2.50	1.50	-	-	-	6.11	7.67	36.88	10.34	49.72
07	(Deslimed)	4/3	35.3	1.7	11.0	2.50	1.50	-	-	-	6.05	7.87	38.61	10.76	52.79
14	Incr. silicate	4/3	35.3	20.0	6.0	2.50	3.50	-	-	0.020	5.37	6.92	26.23	11.27	42.72
											(4.90	7.39	32.79	11.27	49.99)
15	Incr. Oleic/time	4/3	35.3	23.8	10.5	3.50	3.50	-	-	0.012	3.63	5.64	32.61	9.79	56.62
											(3.46	6.17	42.63	9.79	67.95)
16	Decr. Oleic	4/3	35.3	21.3	9.5	1.72	3.50			0.012	5.59	6.91	22.07	11.59	36.99
											(5.97	7.55	28.07	11.59	43.10)
09	pH 2	4	16.0	-	2.0	1.00	-	-	64.97	0.030	7.61	7.58	1.85	6.25	1.52
04	pH 3-5	4	16.0	-	9.0	1.50	-	-	33.75	-	7.22	7.28	8.42	7.90	9.14
05	pH 7-8	4	16.0	-	6.0	1.50	-	-	-	-	7.58	7.68	22.54	8.04	23.57
08	pH 10	4	16.0	-	8.0	1.00	-	0.90-		-	7.22	7.53	29.84	8.24	32.68
30	SO_2 pretreat	15	86.3	21.9	8.5	4.00	3.05	0.26	4.90(4)	0.054	7.20	7.98	27.34	10.04	34.40
											(7.73	8.54	35.04	10.04	41.19)
33	Further upgrading	4/3	66.0	7.3	13.5	4.00	3.00	-	-	0.027	5.36	7.12	26.30	12.06	44.52 (3)
											(5.56	7.40	28.40	12.06	46.22)

Feed - P1 ore for Test 1-16; P-02 ore in Tests 30 and 33

1) - tail assay includes flotation tails and slimes, if desliming was performed
2) - bracketed numbers - flotation stage only
3) - subsequent reverse flotation produced product with 12.64% Mn, 24.98% Mn in 14.07% of weight
4) - 4.90 kg/t SO_2 in a pretreatment stage.

4.6.1.3 Effect of Sulphidization

The deslimed ore in Test MNF 30 was sulphidized with SO_2 prior to flotation. Improved recovery or grade did not result.

4.6.1.4 Direct Followed by Reverse Flotation

The flotation concentrate of Test MNF 33 was reverse floated after oxidation with hypochlorite. The reverse flotation increased product grade from 10.0 to 12.6% Mn and overall recovery was decreased from 34.4 to 2.49%. Therefore, staged flotation did not significantly improve product grade.

Based upon the preliminary direct flotation data, more emphasis was put on reverse flotation testwork.

4.6.2 Reverse Flotation of Gangue Minerals from the Ore

The results of the reverse flotation testwork are summarized in Table 12. The initial five tests investigated the flotation of iron oxides, calcite and silica with Aero Float 899/fuel oil, Aero Float 845/fuel oil, and Armac C collector/dextrin depressant respectively.

The Armac C amine reagent was successful in providing significant upgrading in Test MNF 03. Further testwork was therefore conducted with the amine.

4.6.2.1 Effect of Grind

The grind was varied in Tests MNF 24-28. Desliming was not performed. The results are summarized in Figures 1 and 2. Increased grinding resulted in some increase in product grade up to 85% -200 mesh. Finer grinding also resulted in decreased manganese recovery.

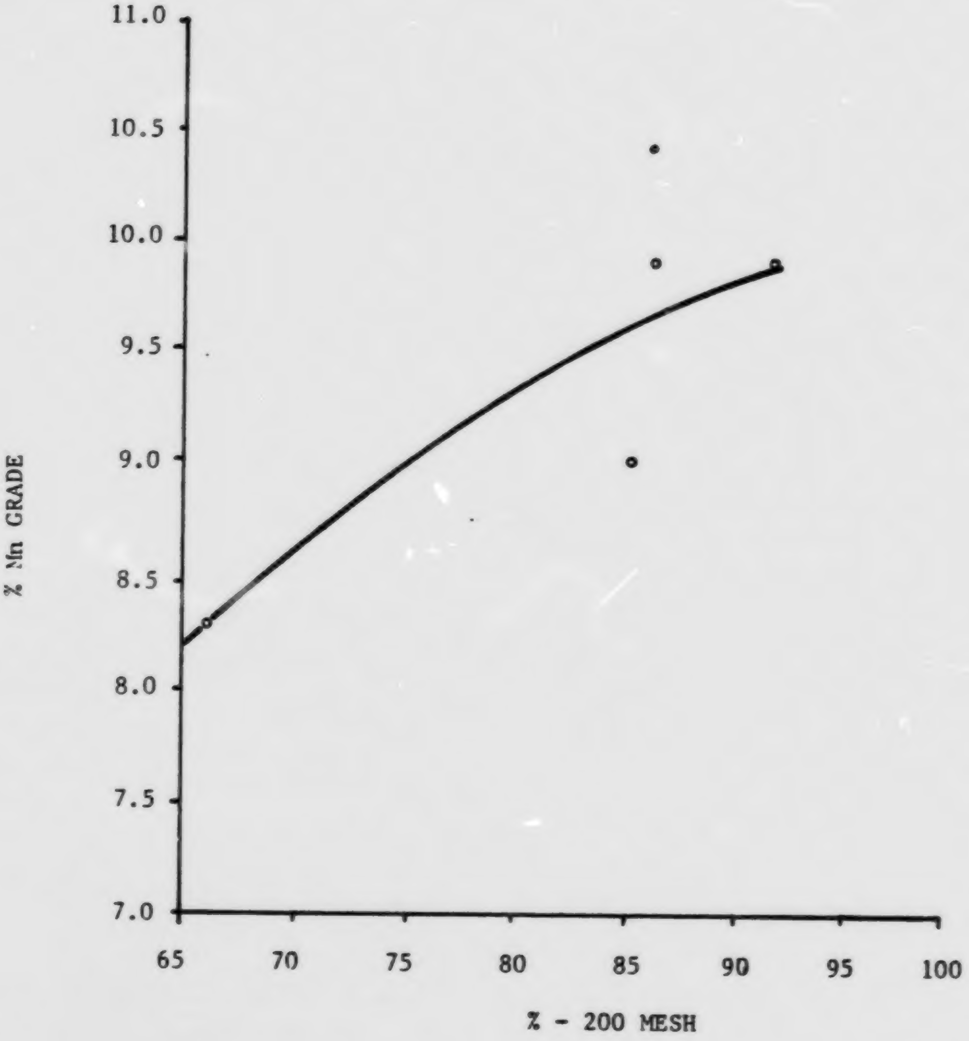
Table 12

SUMMARY OF REVERSE ORE FLOTATION TESTS (9)

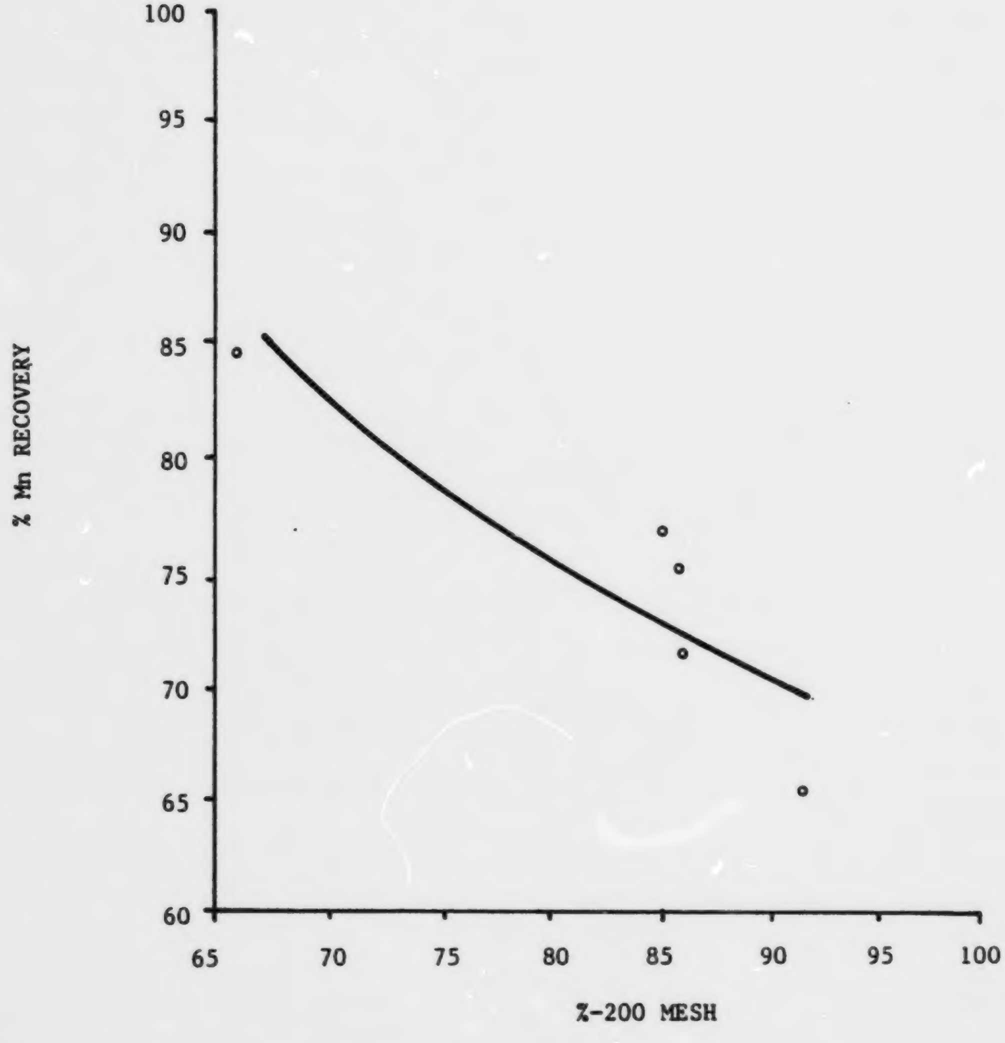
Test #	Purpose		Grinding		Wt % Deslimed	Flot'n Time (min)	Reagents (kg/t)						Float & Slimes (% Mn)	Calc. Head (%Mn)	Flotation Tail		
	Mineral	Collector	min.	%-200 Mesh			Collector	Fuel	NaOH	HMP	Frother	Dextrin			% Wt.	%Mn	%Mn Dist
MNF																	
11	FeOxides	899/Fuel	4	16.0	-	7.0	2.00	2.60	- 35.62(8)	-	-	-	7.06	8.66	89.66	8.84	91.57
10	Calcite	845/Fuel	4	16.0	-	11.0	1.50	1.14	105.05(8)	-	0.044	-	4.33	7.58	94.18	7.78	96.68
12	Calcite	845/Fuel	4	16.0	-	7.0	1.24	1.14	-	-	0.024	-	7.45	8.01	88.48	8.08	89.28
13	Calcite	845/Fuel	4	16.0	-	11.0	1.50	1.14	6.42(1)		0.020	-	7.72	7.44	75.31	7.35	74.40
03	Silica	Armac C	4	16.0	-	10.0	0.70	-	1.16	-	-	0.75	2.72	8.29	83.06	9.43	94.45
17	Silica	Armac C	4/3	35.3	22.7	11.0	1.26	-	1.16	-	0.054	5.00	4.00	7.18	33.66	13.45	63.04
													(3.30	7.72	43.56	13.45	75.88)(3)
18	Silica	Armac C	4/3	35.3	25.3	11.5	1.10	-	1.72	-	0.027	5.00(2)	4.95	6.78	65.44	7.75	74.76
													(4.61	7.36	87.64	7.75	92.26)(3)
19	Silica	Armac C	4/3	35.3	22.5	11.0	1.00	-	1.24	-	0.027	5.00	4.32	6.89	36.98	11.26	60.47
													(3.72	7.32	47.74	11.26	73.44)(3)
22	Deslime	Armac C	15	86.3	32.4	11.0	1.18	-	1.12	1.00	0.054	4.60	4.91	8.06	31.58	14.88	58.29
													(4.63	9.42	46.71	14.88	73.80)(3)
23	Deslime	Armac C	15	86.3	3.6	11.0	1.42	-	1.88	1.00	0.027	5.60	5.79	8.54	49.70	11.32	65.89
													(5.97	8.73	51.54	11.32	66.86)(3)
27	No Desliming		15	86.3	-	11.0	1.58	-	1.92	-	0.027	6.20	4.60	7.95	57.02	10.48	75.14
	Grind,Time	Armac C				15.5	1.78	-	2.16	-	0.027	6.20	4.41	7.95	40.88	13.08	67.21
25	Grind	Armac C	4/3	66.0	-	11.0	1.58	-	1.88	-	0.027	6.20	2.89	6.47	65.79	8.33	84.7
24	Grind	Armac C	10	85.5	-	11.0	1.58	-	1.88	-	0.027	6.20	3.73	6.83	58.69	9.01	77.43
26	Grind	Armac C	20	91.9	-	11.0	1.58	-	2.12	-	0.027	6.20	5.28	7.60	49.90	9.93	65.20
28	As MNF27	Armac C	15	86.3	-	11.0	1.58	-	2.24	-	0.013	6.20	5.08	7.78	56.46	9.86	71.54(5)4
29	Deslime (As MNF 25)	Armac C	4/3	66.0	18.4	11.0	1.25	-	1.35	1.00	0.027	5.00	3.28	6.97	37.00	13.26	70.35(10)
													(2.50	7.38	45.34	13.26	81.84)(3)
31	SO_2 pretreat	Armac C	15	86.3	5.5	15.0	1.43	5.10(7)	7.37	1.00	0.027	5.00	3.69	7.25	44.09	11.76	71.55
													(3.64	7.43	46.63	11.76	73.84)(3)
32	As MNF29	Armac C	4/3	66.0	7.1	15.0	1.24		1.87	1.00	0.027	5.00	3.67	7.12	50.07	10.57	74.27(6)4
													(3.63	7.37	53.90	10.57	77.29)(3)

1) - Sodium carbonate
2) - Quebracho
3) - Bracketed numbers flotation stage only
4) - Further upgrading by oleic flotation
5) - Subsequent direct flotation produced conc. with 11.20% Mn, 19.18% of Mn in 13.32% of wt., subsequent cleaning with oleic provided concentrate with 12.29% Mn, 11.82% of Mn in 7.48% of wt.
6) - Subsequent direct flotation produced conc. with 12.53% Mn, 6.10% of Mn in 3.59% of wt.
7) - 5.10 kg/t SO_2 in pretreatment stage
8) - HCl
9) - Feed for tests 22-32 was P-02 composite; feed for other tests was ore P1.
10) - As MNF 25

EFFECT OF GRIND ON GRADE IN REVERSE FLOTATION



EFFECT OF GRIND ON RECOVERY IN REVERSE FLOTATION



4.6.2.2 Effect of Flotation Time

Flotation time was varied in Tests MNF 27 and 28. Product grade after 11 minutes of flotation was 9.8-10.4% Mn, and recovery 70-75%. After an additional 4.5 minutes, grade had increased to 13.0% and recovery was 57%.

4.6.2.3 Effect of Desliming

Desliming seemed to result in increased flotation product grades. Slimes representing 0, 3.6, 5.5, 18.4 and 32.4% of the weight were removed in Tests MNF 27, 23, 31, 29 and 22 respectively for 86% -200 mesh grind products. Curves illustrating product grade and recovery versus % deslimed are shown in Figure 3 and Figure 4 respectively.

Incomplete desliming was somewhat compensated for by additional flotation time. Therefore, similar grades and recoveries were achieved by either process.

Figure 3

EFFECT OF DESLIMING ON PRODUCT GRADE IN REVERSE FLOTATION

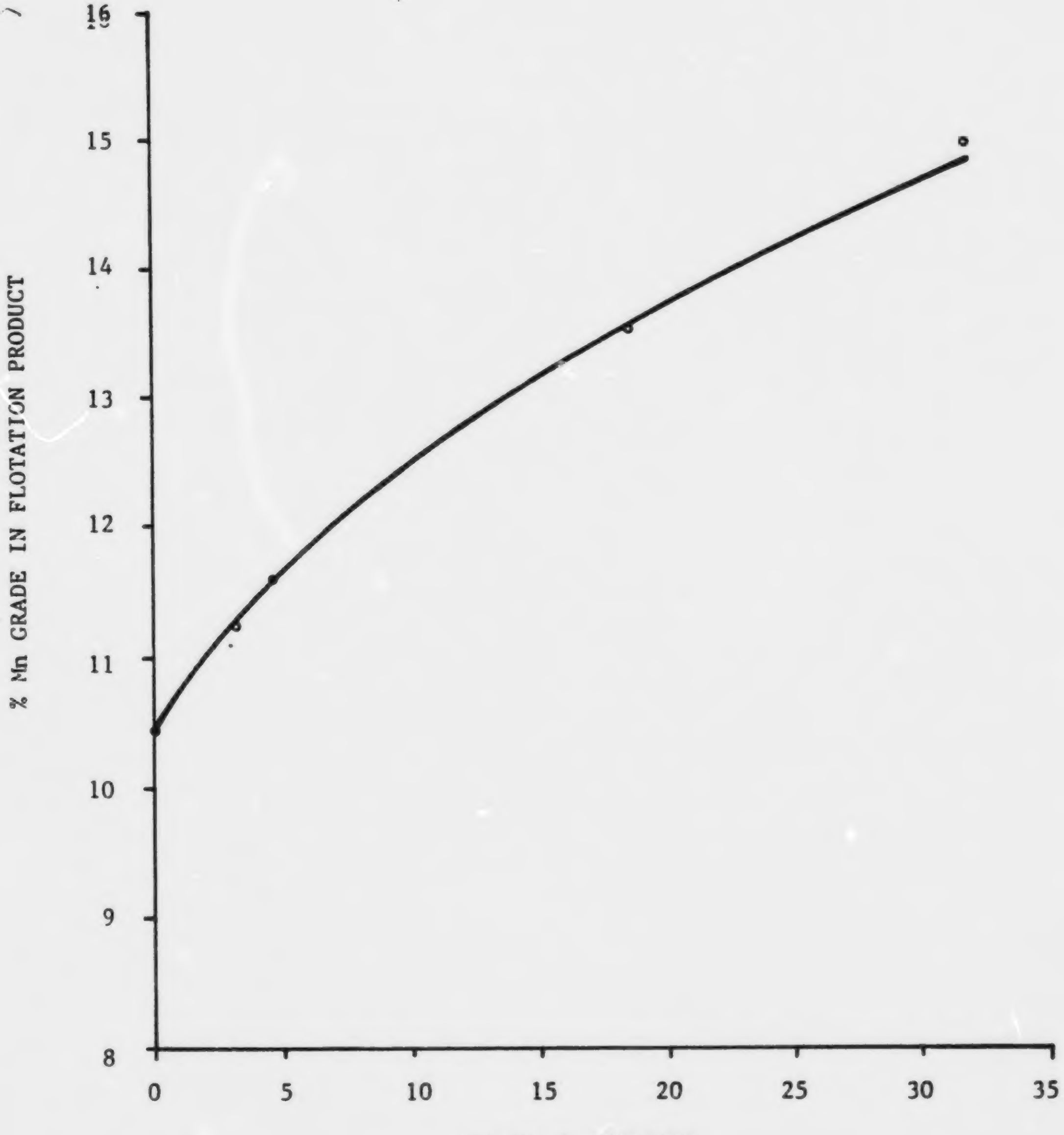
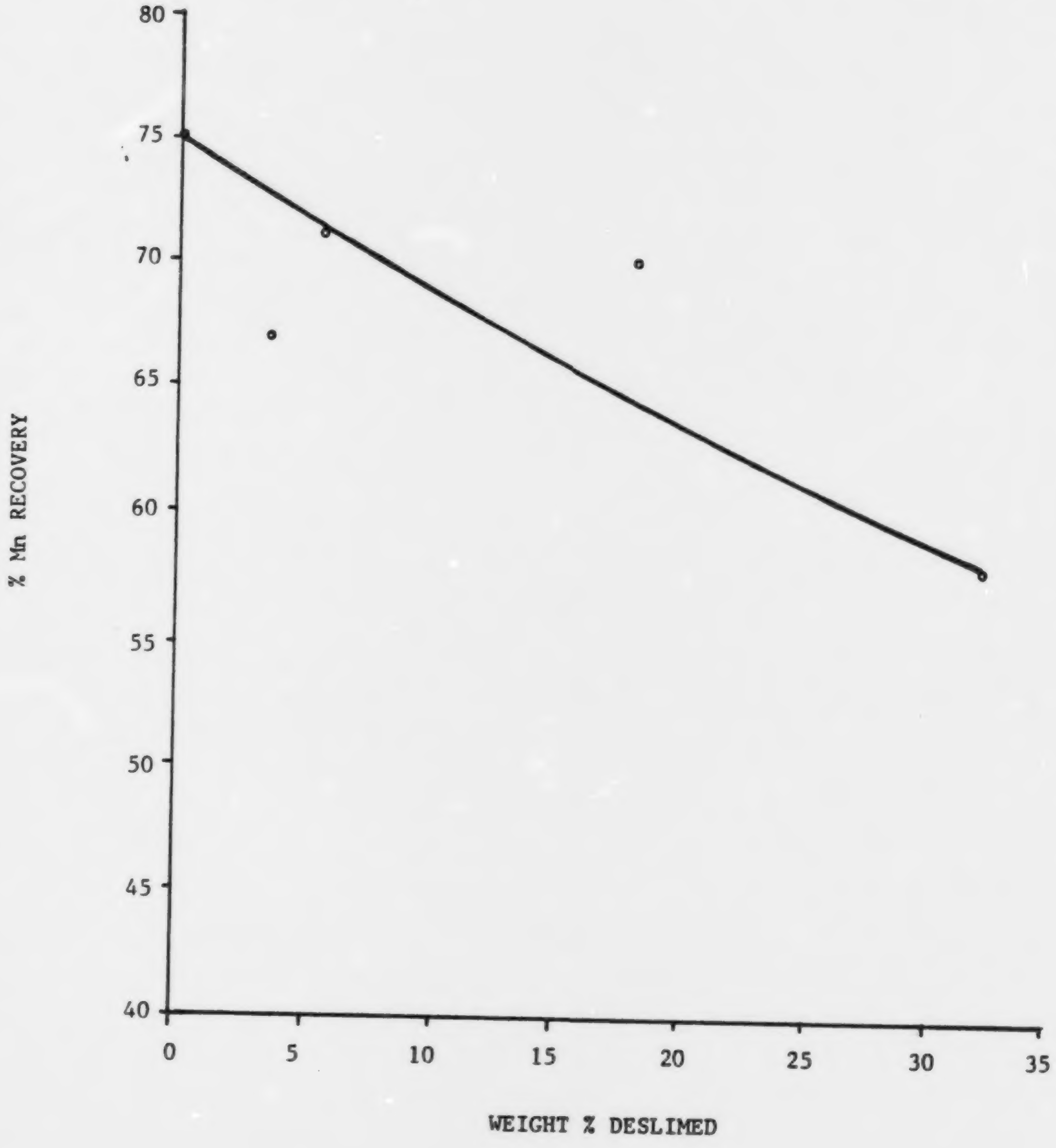


Figure 4

EFFECT OF DESLIMING ON RECOVERY
IN REVERSE FLOTATION



4.6.2.4 Effect of Sulphidization

The deslimed ore was sulphidized with SO_2 prior to flotation in Test MNF 31. Improved flotation results were not obtained.

4.6.2.5 Reverse Followed by Direct Flotation from the Ore

Reverse flotation was followed by direct Mn flotation with oleic acid in Tests MNF 28 and 32. Upgrading from the reverse float tail was extremely limited, as grades were raised from 9.8% Mn to 11.2% Mn in Test 28, and from 10.5% Mn to 12.5% in Test 32. In addition, subsequent cleaning stages with oleic acid were unsuccessful in raising the product grades above 12-13% Mn. Oleic acid flotation stage recoveries were very poor, so that overall manganese recoveries were less than 20%.

4.6.3 Composition of Flotation Product

A semi-quantitative analysis was performed on a composite of the products of reverse and direct flotation. The results are summarized in Table 13. There was some upgrading of Mn, Fe, P, Ba, Sr, As, Mo, and Au.

4.6.4 Flotation of Spiral Products

The products of spiral Test MSP 14 were floated to upgrade the manganese levels. The feed materials were ground, deslimed, reverse floated, and then directly floated. Test parameters were derived from optimum ore flotation conditions. The following samples were tested:

1) MSP 14 concentrate
2) Combined MSP 14 concentrate and middling
3) -65 mesh ore (screened out prior to spiral separation of the +65 mesh fraction).

Table 13

COMPOSITION OF FLOTATION PRODUCTS COMPOSITE *

	Wt %			ppm			ppm	
Element	Feed	Products	Element	Feed	Products	Element	Feed	Products
SiO_2	38.7	22.0	Ba	1830	2580 **	Th	7.4	4.4
Fe_2O_3	14.8	19.6 **	Sr	260	380 **	Be	6.0	5.0
Al_2O_3	11.0	5.62	Cr	120	82	Sm	5.4	6.3
MnO	9.69	17.3**	Rb	100	10	Cs	4.3	2.5
CaO	6.65	11.5	Ni	78	63	Pb	4.0	2.0
K_2O	2.59	1.18						
MgO	2.50	1.75	Co	77	89	Hf	3.1	1.9
Na_2O	0.91	0.59	As	77	140 **	W	3.0	4.0
			Zr	70	30	Yb	2.23	2.54
P_2O_5	0.55	1.10 **	Ce	70	71	U	2.20	1.9
TiO_2	0.49	0.28	Cu	46	57	Mo	2.0	9.0 **
			Li	46	34	Br	0.8	0.8
			B	40	20	Se	1.4	0.7
			La	31.1	29.9	Ev	1.26	1.76
			V	28	12	Bi	0.9	0.40
			Nd	25	25	Tb	0.8	1.0
			Nb	20	20	Lu	0.34	0.36
			Y	20	20	Ta	LT 0.5	LT 0.5
			Sc	14.7	9.27	Ag	LT 0.5	LT 0.5
			Ge	LT 10.0	10	Cd	LT 0.2	LT 0.2
			Sb	7.8	6.3	Au	0.073	0.370**

* Plymouth P1 ore

** Some upgrading by flotation

LT Less than

Conditions and results are summarized in Tables 14-16 and Figure 5. Flotation failed to significantly increase product grades. Processing of the ore by gravity and flotation methods to achieve acceptably high grades at the expense of recovery was therefore not successful. Products were readily upgraded by a factor of two, but the degree of liberation for both coarse and very fine ore did not allow for further upgrading.

Further mineral processing testwork was therefore not conducted. If any mineral processing is included in a plant flowsheet, it is expected that it would be limited to a single stage of gravity separation.

4.6.5 Mineralogy of Flotation Products

Examination of the reverse flotation products revealed that manganese carbonate losses were typically low while losses of manganese oxides and silicates were more extensive. Reverse flotation provided cleaner products and higher manganese mineral recoveries than did direct flotation. Liberation and flotation response are the major sources of losses. This confirms the conclusions deduced from the metallurgical results. Photomicrographs of selected flotation products are shown in Appendix VI.

Observations of direct flotation products indicated that recoveries of manganese oxides and carbonates were high, although oxide losses were noted. Manganese silicate recovery was less complete. Carbonate flotation was considerably more rapid than that of the oxides, although this data is based primarily upon the major mineral content of the oberved particles. Manganese losses were definitely due to non-liberation although enhanced by flotation response in terms of flotation time and possibly depression by sodium silicate.

Table 14

CONDITIONS FOR SPIRAL PRODUCT FLOTATION TESTWORK

Test			Grinding		Flot'n Time (min)		Reagents (kg/t)						
No.	Feed	Deslimed	min.	% -200 mesh	Reverse	Direct	Armac C	Oleic	NaOH	HMP	Na Silic	Frother	Dextrin
36	Spiral Conc.	No	4/3	66.0	15.0	5.2	1.58	0.31	1.53	-	1.74	0.054	6.00
37	Spiral Conc.	Yes	4/3	66.0	15.0	5.2	1.22	0.31	1.40	1.00	1.74	0.054	5.00
38	Spiral Conc. + Middling	Yes	4/3	66.0	15.0	5.2	1.22	0.31	1.45	1.00	1.74	0.054	5.00
34	-65 mesh ore	No	1	NA	15.0	5.2	1.58	0.31	1.44	-	1.74	0.054	6.00
35	-65 mesh ore	Yes	1	NA	15.0	5.2	1.22	0.31	1.39	1.00	1.74	0.054	5.00

Table 15

FLOTATION OF SPIRAL PRODUCTS

Test #	Feed Type	Feed % Mn	Reverse Flotation % Mn	Reverse Flotation % Wt.	Reverse Flotation % Rec.	Direct Flotation % Mn	Direct Flotation % Wt.	Direct Flotation % Rec.
36(1)	Spiral Con.	11.17	14.25	76.07	88.69	14.85	15.62	18.98
37	Spiral Con.	11.17	13.93	69.06	81.46	16.05	12.78	17.37(2)
38	Spiral Con. + Middling	9.77	13.83	71.60	85.47	16.41	15.62	22.13
34(1)	-65 mesh Ore	6.65	10.13	44.02	73.17	9.83	5.42	8.74
35	-65 mesh Ore	6.65	11.95	34.20	61.44	13.50	6.60	13.40

1) Not deslimed

2) Subsequent cleaning with oleic acid produced a concentrate with 16.74% Mn, 11.12 % of Mn in 7.85% of weight.

Table 16

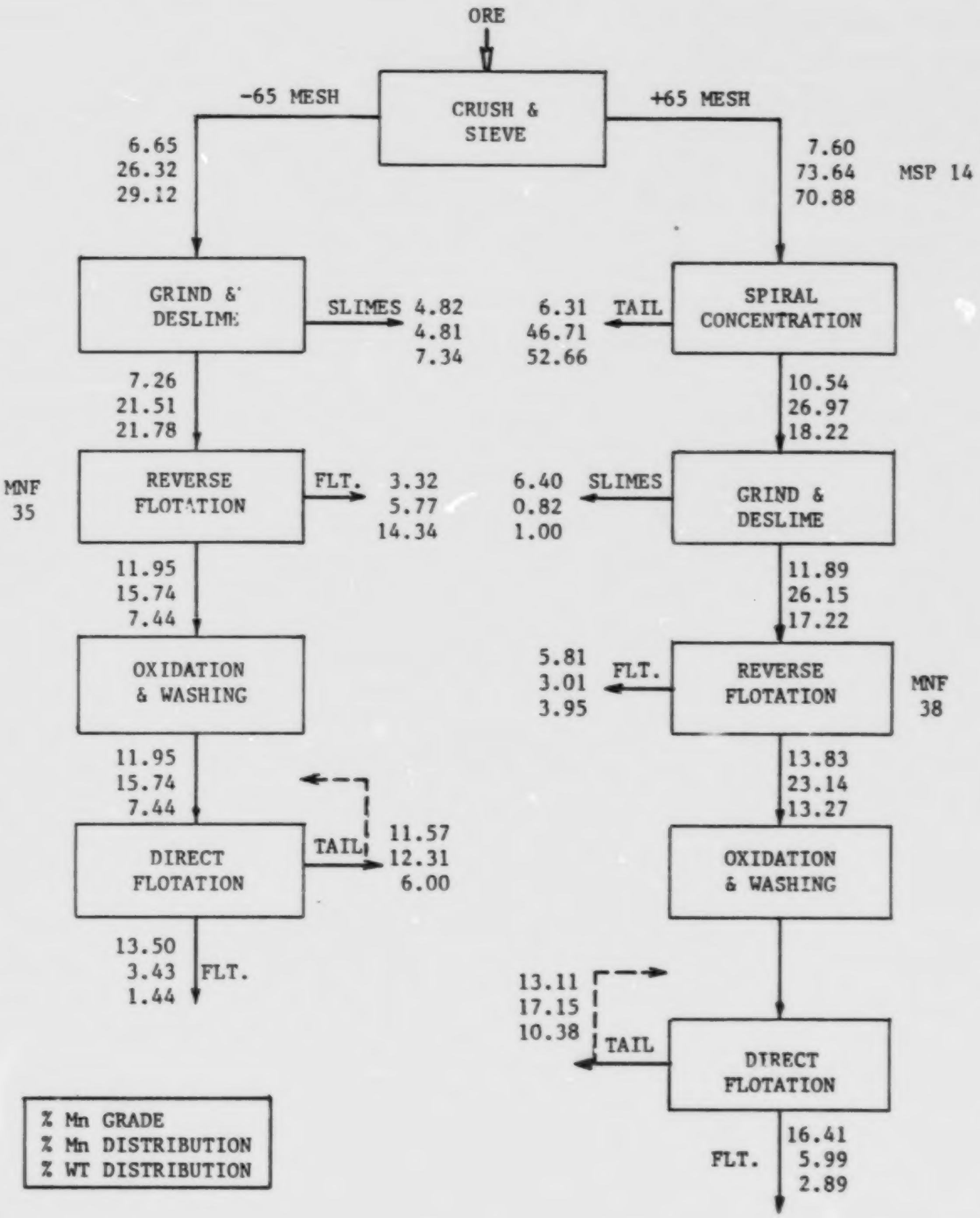
OVERALL RECOVERY BY A GRAVITY AND FLOTATION FLOWSHEET

	Grade % Mn	% Distribution Wt.	% Distribution Mn
Reverse Flotation	13.15	20.71	38.88
Reverse/Direct Flotation	15.44	4.33	9.42

* Based upon Spiral Test MSP 14 Conc. and middling and Flotation Test MSP 35 and 38. See Figure 5.

Figure 5

FLOWSHEET OF
GRAVITY-FLOTATION TESTWORK


ORE
CRUSH & SIEVE
-65 MESH
+65 MESH
6.65
26.32
29.12
7.60
73.64
70.88
MSP 14
GRIND & DESLIME
SLIMES 4.82
4.81
7.34
6.31
46.71
52.66
TAIL
SPIRAL CONCENTRATION
7.26
21.51
21.78
10.54
26.97
18.22
MNF 35
REVERSE FLOTATION
FLT. 3.32
5.77
14.34
6.40
0.82
1.00
SLIMES
GRIND & DESLIME
11.95
15.74
7.44
11.89
26.15
17.22
OXIDATION & WASHING
5.81
3.01
3.95
FLT.
REVERSE FLOTATION
MNF 38
11.95
15.74
7.44
13.83
23.14
13.27
DIRECT FLOTATION
TAIL
11.57
12.31
6.00
OXIDATION & WASHING
13.50
3.43
1.44
FLT.
13.11
17.15
10.38
TAIL
DIRECT FLOTATION
FLT.
16.41
5.99
2.89
% Mn GRADE
% Mn DISTRIBUTION
% WT DISTRIBUTION

Observations made on the flotation of spiral products confirmed these results from direct and reverse flotation of ore samples. Although increased oxide losses were noted, the concentrates contained numerous high Mn-oxide particles.

In summary, flotation successfully recovered the liberated and exposed manganese-rich particles. Losses were primarily due to non-liberation and selective flotation responses by oxides and carbonates. In addition, losses of manganese silicates in the direct flotation may be, in part, due to the flotation reagent scheme.

The mineralogical examinations confirmed the results and observations of the metallurgical testwork and the preliminary mineralogical studies.

4.7 Extractive Metallurgy

Roast-leach and leaching tests were conducted on the Plymouth ore and flotation product in order to extract manganese. Test conditions and mass balances are detailed in Appendix VII.

4.7.1. Flotation Products

Roasting and leaching tests were conducted on a composite of the flotation products of Tests 14-19. The makeup of the composite is shown in Table 17. The assayed head values were 11.52% Mn, 8.18% Fe and 0.008% Y_2O_3. Test results are summarized in Table 18. The calculated Mn assay was 12.34%.

Table 17

COMPOSITION OF FLOTATION PRODUCT COMPOSITE

Test	Product	Wt%	Grade % Mn
MNF 14	Conc. 1	7.8	10.40
	2	10.1	13.89
15	Conc. 2	4.1	16.89
	3	3.0	10.42
16	Conc. 1	5.0	12.30
	2	0.9	13.94
	3	9.6	10.96
17	Tail	28.5	13.45
18	-	-	-
19	Tail	31.0	11.26
Total		100.0	(12.34)

Table 18

MANGANESE EXTRACTION FROM FLOTATION PRODUCTS

Test	Process	Roasting Temp C	Roasting Time (hr)	Leaching Time (min)	Na_2SO_3	H_2SO_4	NaCl	SO_2	Mn ppm	Mn % Extr'n
MEX- A-1	Sulphatizing Roast	400-675	5.5	20.10	504	-	-	-	0.01	tr
A-2	Sulphatizing Roast	400-675	5.5	20.0	240	-	-	-	tr	tr
B	Oxidizing Roast	500	1.0	20.0	-	384	480	-	326	23.73
C	Reducing Leach	-	-	6.0	-	*	-	*	789	71.67

* Additions not measured

Virtually no manganese was dissolved from the sulphatizing roast calcine. Mn extraction from the oxidizing roast calcine was 22.9% after 3 hours of leaching. Leaching for a further 17 hours resulted in an increase in Mn extraction of only 1%.

Manganese extractions in the reducing (SO_2/H_2SO_4) leach were 45.6, 63.0, and 71.7% after 2, 4 and 6 hours respectively. Mn tenor in solution was as high as 0.79 gpl.

Therefore, a reducing leach with H_2SO_4 and SO_2 was successful in extracting a high percentage of the manganese in the float products. Further reduction leach testwork on samples of the Plymouth ore was conducted using these initial data as a starting point.

It should be noted that test parameters were not optimized and reagent levels were higher and pulp densities lower than would be used in practice.

Due to unacceptably low flotation grades, further extractive metallurgy testwork with flotation products was not considered.

4.7.2 Heap Leaching of Ore with SO_2

A coarse (0.95 cm) sample of the Plymouth P-02 ore was leached with SO_2/H_2SO_4 for 49 days at a pH of less than 2. Results are summarized in Table 19 and Figure 6.

Mn extraction was 71% after 49 days. Leaching was essentially complete after 33 days. Maximum Mn tenor in solution was 3 gpl. Iron extraction was 19.5%. The Fe tenor in solution was 1.5 gpl.

Table 19

SUMMARY OF SO2 LEACHING OF PLYMOUTH ORE SAMPLES

Test #	Target pH	Time (days)	Additions H_2SO_4 (kg/t)	Additions SO_2 (kg/t)	Additions SO_2 (%of Stoich)	Residue (%Mn)	Mn in Sol'n (gpl)	Calc. Hd (% Mn)	% Mn Extr'n	Details
MHL 1	3/<2	49	213.2	526.4	603	2.72	3	7.32	71.59	1)
MBL 1	1.5	13	127.8	-	-	5.54	16	6.67	26.95	2,3
MBL 2	2.5	14	149.2	299.9	345	3.80	16-18	7.63	59.13	2
MBL 3	2.5	14	162.9	750.2	862	3.46	12-16	6.41	57.03	2
MBL 4	1.5	5	212.0	375.1	431	NA	32	NA	42.48	2
MBL 5	2.0	10	179.0	90.0	103	1.88	27-7	8.20	80.50	2,4
MBL 6	2.0	10	116.4	90.0	103	3.90	12-5	7.89	57.40	2
MBL 7	3.0	10	54.0	90.0	103	4.77	9-3	7.19	39.13	2
MBL 8	2.0	10	107.6	90.0	103	4.05	32	7.55	52.14	2,3

1) Heap Leach in 5 cm diameter column; SO_2 gas used

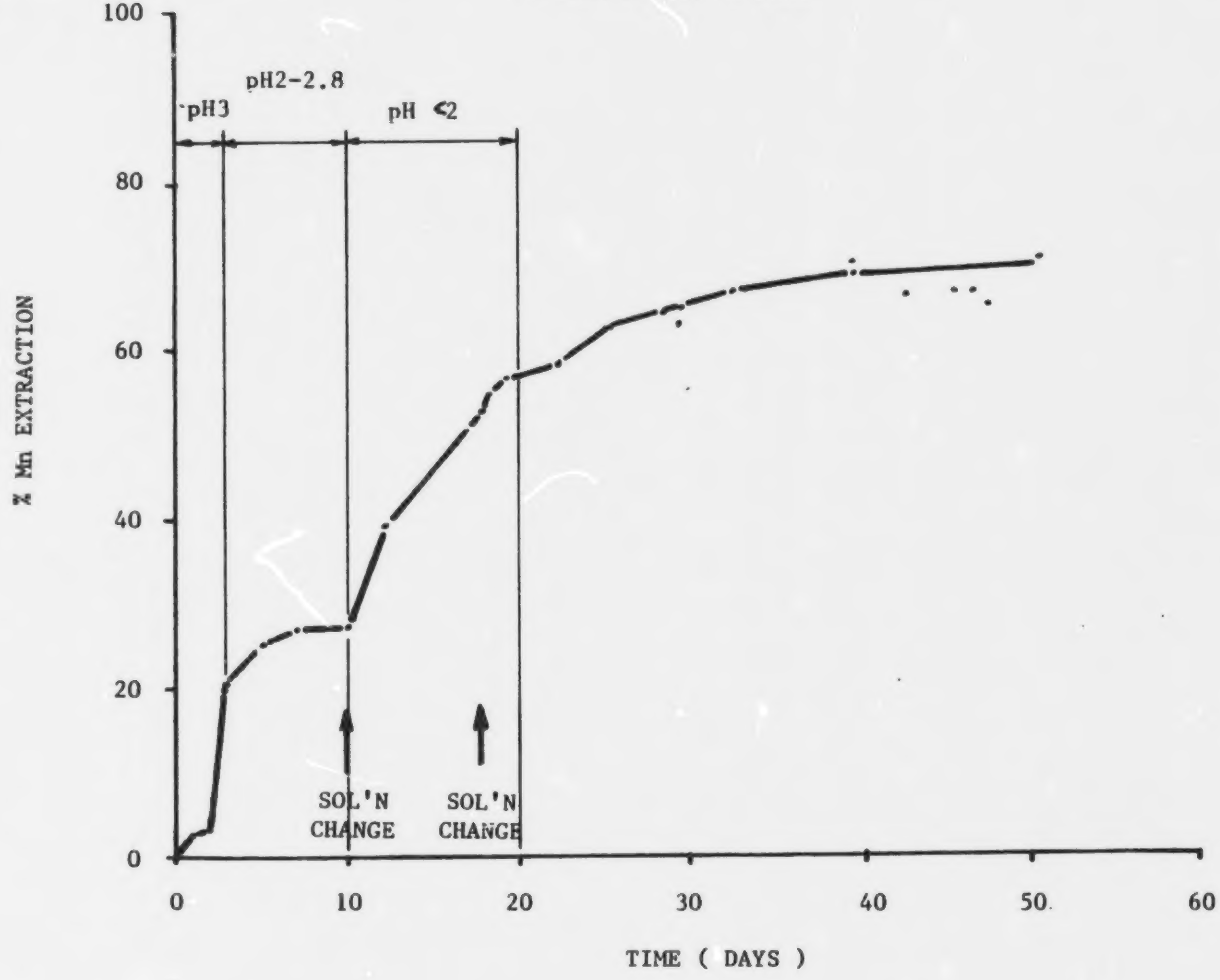
2) Sodium metabisulphite used as SO_2 source; additions converted to equivalent SO_2 levels
Leaches in bottles at 28 rpm; capped

3) Solution not changed during test; all other tests had solution changed 1-4 times

4) Feed ground to approximately 85% -200 mesh

Figure 6

COLUMN HEAP LEACH OF PLYMOUTH ORE
Mn EXTRACTION VS TIME



Sulphur dioxide consumption was extremely high. This was primarily due to the method of addition, and is not representative of expected consumptions. Based upon these data, further tests were conducted in bottles to determine the effect of leach parameters such as size, pH, SO_2 level, sulphuric acid level and repulping.

4.7.3 Bottle Leaching of Ore with Sodium Metabisulphite

Eight bottle rolls tests were conducted on the minus 0.95 cm ore with $Na_2SO_2O_5$ and H_2SO_4. The results are summarized in Table 19.

4.7.3.1 Tests 1-4

Tests MBL 1-4 were conducted to determine the effect of sulphuric acid, pH, and SO_2 level. The results can be summarized as follows:

- The very high SO_2 and H_2SO_4 additions in Test MBL4 resulted in uncontrollable fizzing.
- Sulphuric acid alone in MBL 1 extracted 27% of the manganese, presumably due to carbonate leaching.
- Excessively high SO_2 addition did not result in increased Mn extraction.
- Maximum Mn extraction was 59% in 14 days. This correlates well with 41% Mn extraction after 14 days in the heap leach.

Table 20

ANALYSIS OF SO_2 LEACH SOLUTION

Element	Tenor (ppm)	Element	Tenor (ppm)
S	"High"		
Na**	13,400	Te	7.5
Mn	10,400	Sr	7.2
Fe	5,800	Hg	7.0
Al	1,100	Cr	5.8
Si	575	Sn	5.6
Ca	537		
Mg	475	Zn	5.5
K	215	W	5.1
P	123	Ni	4.9
Pb	100	Mo	2.5
C	85	Ag	3.1
Nb	44	V	2.2
		Ce	1.7
U	20	Y	1.1
Se	18.5	Zr	0.9
Sb	16.8	Cd	0.65
B	12.2		
Bi	11.2	Ti	0.5
Th	8.4	Be	0.23
Cu	7.9	Ba	0.06
Co	7.6	As	0.005

* Semi-quantitative analysis of MBL8 pregnant solution.

** High due to use of sodium metabisulphite

4.7.3.2 Tests 5-8

Further bottle leaching tests were conducted to determine the effect of particle size, pH, the effect of solution change, and SO_2 level. Retention time was 10 days for each test. Change of solution was performed every two days, except for Test 8. The results are presented in Table 19 and Figure 7. The results may be summarized as follows:

- grinding to 85% -200 mesh in Test 5 resulted in a manganese extraction of 80%. Extraction for the crushed (0.95cm) ore was 57%. These levels were achieved with 100% stoichiometric SO_2 (ie- SO_2 requirement for dissolution of all Mn. In practice, Mn carbonate would not be leached by SO_2).
- change-out of solution resulted in a 5% increase in Mn extraction; the effect was noted at a Mn tenor of 22 gpL.
- extraction at pH 2 was significantly higher than at pH 3, indicating that H_2SO_4 level is important.
- Fe dissolution was only 2-3%
- A scan of the pregnant solution of MBL8, shown in Table 20, indicated that the solution composition was complex. This stresses the need for purification investigations during future testowrk.

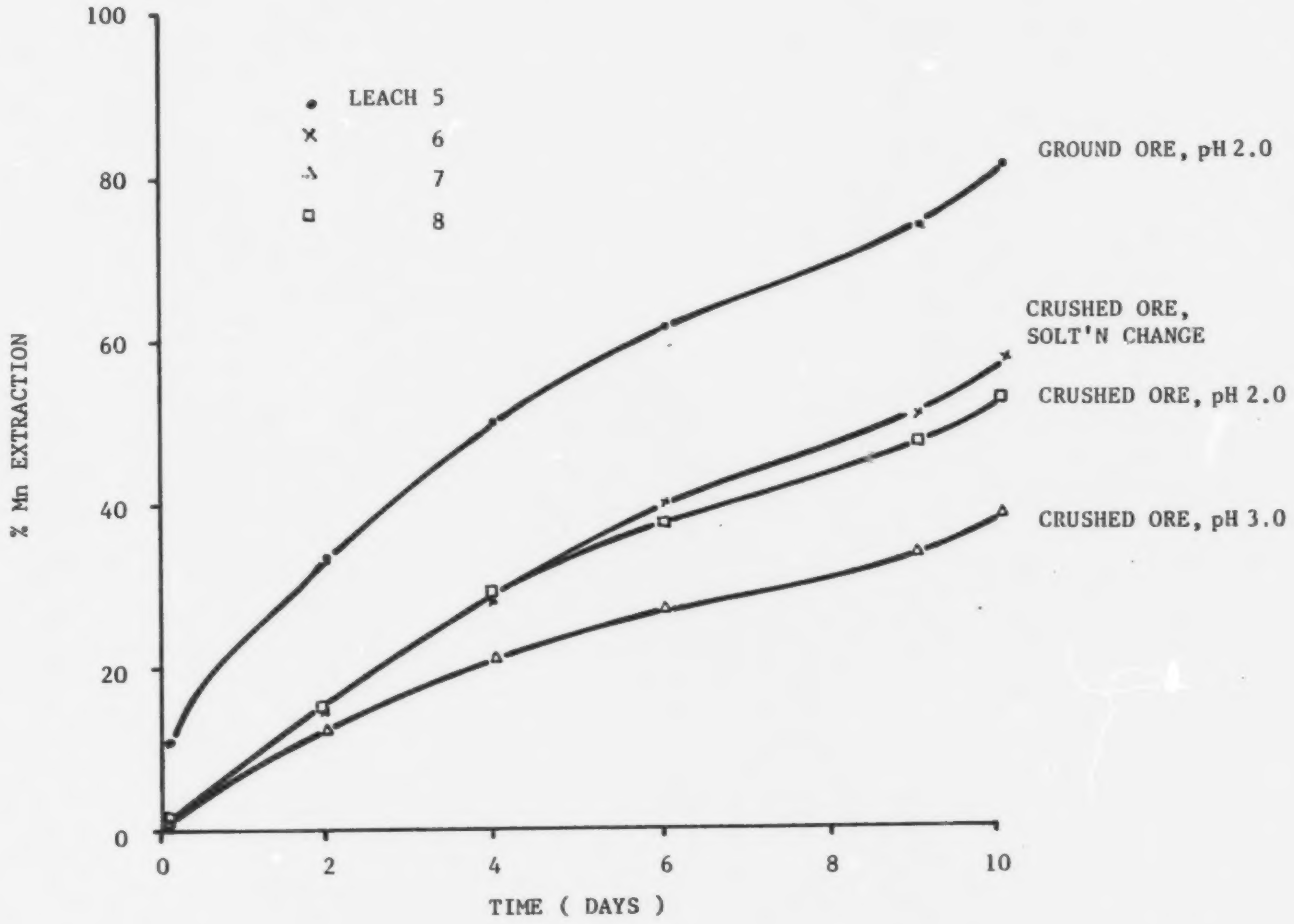
In summary, leaching of the Plymouth ore with SO_2 and sulphuric acid was successful. Reagent requirements were roughly determined. However, it is apparent that heap leaching will dictate high SO_2 consumptions as leaching over 20-50 days will presumably be required.

Further testwork was not planned. It was felt that the remaining time on the project would be most effectively spent in preparing cost studies of various flowsheet options.

Figure 7

SULPHUR DIOXIDE BOTTLE LEACHING OF PLYMOUTH ORE

SERIES B



5.0 CONCLUSIONS AND RECOMMENDATIONS

Based on the results of the testwork, the following conclusions and recommendations are made.

1. Leaching of the coarse ore (-0.95 cm) with sulphuric acid and sulphur dioxide provided manganese extractions of 60 - 71% in approximately 30 days. Acid and SO_2 consumptions were greater than 100 kg/t.

2. Due to the high reagent consumptions, a pre-concentration step is desirable to reduce the level of reactive gangue. In the testwork, all of flotation, spiral concentration and heavy liquid separation were fairly successful as pre-concentration methods. Flotation would of course not be viable as a pre-concentration step for heap leaching. Future testwork should involve examination of heavy media separation as a possible pre-concentration method.

3. Agitated leaching of a flotation concentrate provided an extraction of 71% in 6 hours. Leaching of ground ore in a similar manner yielded an extraction of 80% in 10 days. Therefore, leaching kinetics for the concentrate were more rapid. Also, reagent consumptions were considerably lower due to the reduction in gangue content.

4. Leach solutions contained high levels of impurities. No testwork was conducted on solution purification. Further work is required to confirm that standard purification methods will produce an acceptably clean product solution.

5. Agitated leaching of the concentrate after sulphation roasting yielded virtually no soluble manganese. Extraction of manganese by leaching of an oxidizing roast calcine provided a manganese extraction of less than 25%.

6. Mineral processing of the ore was not successful in upgrading the ore to the desired manganese grade. The results may be summarized as follows:

   i. The ore grade was increased by a factor of approximately two by most of the methods tested - gravity (heavy liquid, spiralling) flotation, and high intensity magnetic separation.

   ii. Upgrading of the gravity concentrate was attempted by flotation. No significant upgrading was achieved. This was due to the intimate association of gangue and manganese minerals and the resulting poor liberation.

   iii Maximum single stage reverse flotation recoveries were on the order of 60 - 75% in 30 - 50% of the weight. Slightly higher manganese recoveries were achieved by flotation of the spiral concentrate. By comparison, the best upgrading by spiralling provided a recovery of 52% of the manganese in 27% of the weight. Therefore, while flotation provided higher recoveries, it could only be used as a pre-concentration method prior to leaching.

   iv Heavy liquid separations on coarsely ground ore provided the highest upgrading (17.5% Mn) with a recovery of 70% of the Mn in 26% of the weight. No heavy media processing or heavy liquid testing on coarse (ie. 0.95 cm or larger) was performed. Future testwork should examine these areas.

v High intensity magnetic separation provided similar grade, recovery and weight rejection as flotation. No cleaning was performed. Low intensity magnetic separation did not provide any upgrading.

vi Hand sorting or ore sorting was not performed and was not considered to be worthwhile based upon a visual examination of the ore.

vii The above conclusions on mineral processing testwork were supported by mineralogical examination of selected specimens.

6. Future testwork should focus on the following:

   i Methods of pre-concentration, namely heavy media and spiral separation.

   ii Large scale heap leach tests to improve the definition of reagent requirements and leaching kinetics.

   iii Solution purification (including solvent extraction).

   iv Manganese recovery (by precipitation, crystallization).

APPENDIX I

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APPENDIX II

Bond Work Index

BOND WORK INDEX

Sample: Plymouth 1 Ore (Sample P1)

Screen Size: 106 microns (150 mesh)

Wt. of 700 mls: 1200.0 g

Desired wt. of Undersize: 342.9 g

Cycle	Number of Revolutions	Grams of Undersize	Gbp (gm/rev)
1	100	174	1.74
2	197	286	1.45
3	237	394	1.66
4	206	280	1.36
5	252	334	1.33
6	258	317	1.23
7	279	340	1.22
8	281	332	1.181
9	290	341	1.176
10	292	282	0.966
11	355	353	0.994
12	345	393	1.139
13	301	390	1.296
14	265	324	1.223

Average of final 3 Gbp vales: 1.201 gm/rev.

$$Wi = \frac{49.1}{(P1)^{0.23} \times (Gpb)^{0.82} \times \left(\frac{10}{\sqrt{P}} - \frac{10}{\sqrt{F}}\right)}$$

Where: P1 = Screen Size (microns) = 106
Gbp = Undersize (gm/rev) = 1.201
P = Product 80% passing size (microns) = 80
F = Feed 80% passing size (microns) = 2250
Wi = Work Index (kwh per tonne) = 15.93
W = Work Input (kwh per tonne) = 14.46

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PARTICLE SIZE DISTRIBUTIONS OF BOND TEST FEED AND PRODUCT

The Tyler Standard Screen Scale

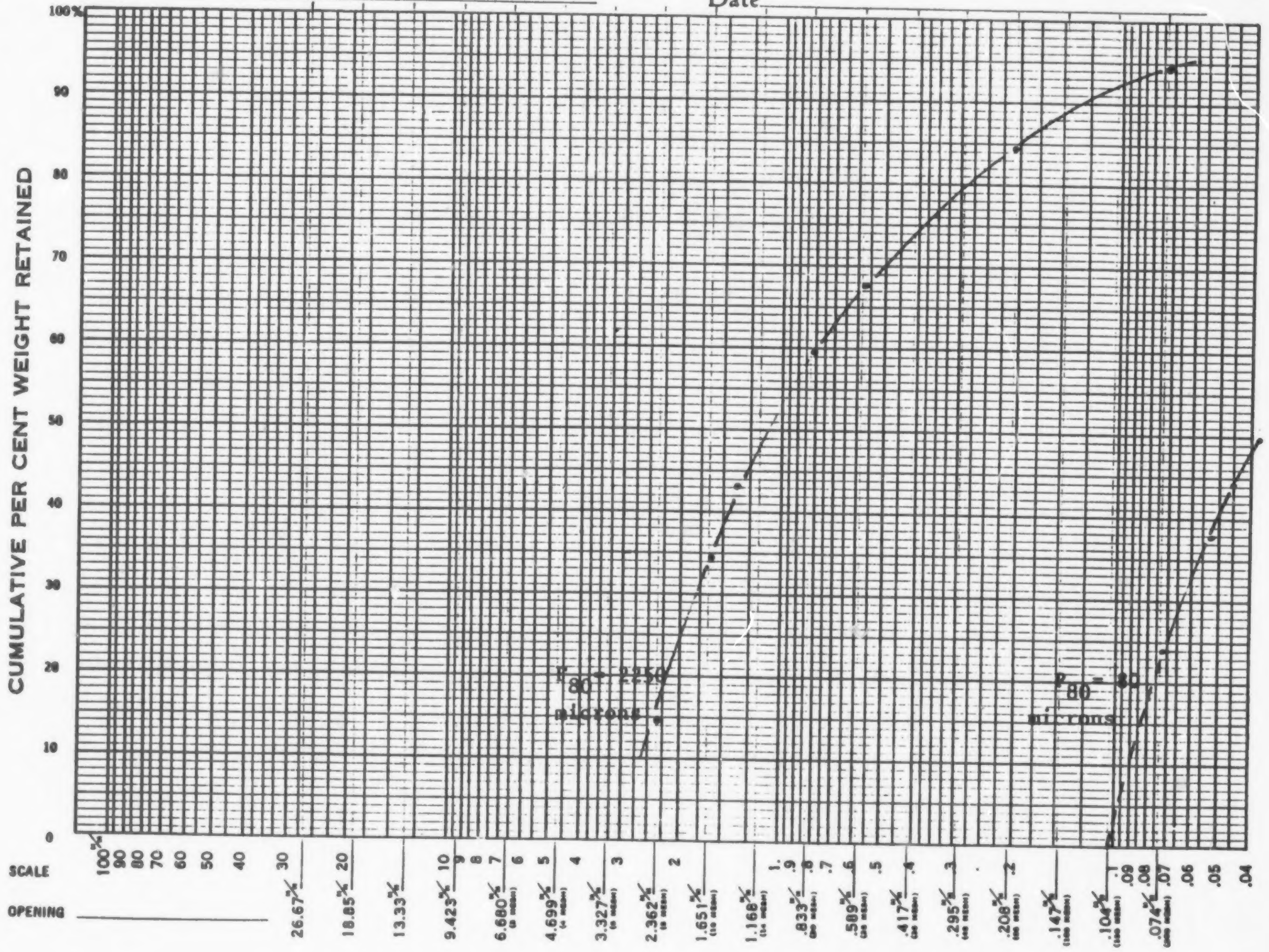
Form No. L-6
Please mention above
when ordering

Cumulative Logarithmic Diagram of Screen Analysis on Sample of P1 ORE

Name

Date

L-6



APPENDIX III

Summary of Screen Analysis Data

PARTICLE SIZE DISTRIBUTIONS OF PLYMOUTH P1 ORE
CRUSHED IN CONE CRUSHER
CUMULATIVE % PASSING

Size Mesh	P1 6 mesh	(1) P1 20 mesh	(2) P-02 20 mesh	(3) P-02 -20 +65 mesh	(4) P-02 -20 mesh	(5) P-02 -65 mesh
3	43.90					
4	29.60					
6	20.10					
8	14.10					
10	10.60					
14		99.80				
20		97.41	99.63	99.67	99.46	
28		80.97	85.12	75.45	78.88	
32			75.52	59.61	65.25	
35	4.90	65.22				
48		49.06	51.36	19.79	30.98	
65	2.90	37.53	39.36		13.95	100.00
100	2.50	18.98	25.64		9.10	65.15
150	2.10	17.00	19.26		6.84	48.93
200	1.70	5.19	14.06		5.01	35.71
Pan	0.00	0.00	0.00	0.00	0.00	0.00

(1) Nominal -20 mesh - used in spiral tests MSP 6-7, and flotation tests MNF 01-19
(2) -20 mesh - used in flotation tests MNF 22-23
(3) -65 mesh of -20 mesh removed - used in spiral tests MSP 8-12
(4) One quarter of -65 mesh returned - used in spiral tests MSP 13-14
(5) -65 mesh - used in flotation tests MNF 34-35

SUMMARY OF PARTICLE SIZE DISTRIBUTIONS OF PLYMOUTH P1 ORE
GROUND IN 35 cm x 20 cm DIAMETER MILL
CUMULATIVE % PASSING

Size (Mesh)	(1) 2 min.	(2) 2/2 min	(3) 3/5 min	4 min	(4) 4/3 min	(5) 5 min	(6) 5 min	10 min
3								
4								
6	100.00					100.00	100.00	100.00
8								
10								
14	66.55		100.00	100.00	100.00	91.60	85.94	100.00
20								
28	45.78		98.81	93.25	92.58	68.93	62.75	95.60
32								
35		73.35						
48	32.40							
65	23.78	41.67	58.83	51.92	47.17	45.26	39.62	64.24
100	23.50	30.29	50.46	42.97	37.33	39.74	34.63	57.16
150		21.17						
200		16.36	37.33	27.43	20.26	29.85	25.05	43.14
270								
325			31.55	20.46	12.60	24.75	21.54	36.81
Pan	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
% Solids	80%	65%	65%	65%	65%	65%	65%	65%

(1) Used in heavy liquid testwork with Rods
(2) +28 mesh reground in stages of 2 minutes - used in spiral tests MSP 01-05 - with balls
(3) Reground +28 mesh for 5 minutes - with rods
(4) Reground -28 +325 mesh for 3 minutes - with rods
(5) Used in Table tests - with rods
(6) Less rods than in other grinds - with rods

PARTICLE SIZE DISTRIBUTION OF PLYMOUTH P1 ORE GROUND IN 20 cm x 20 cm DIAMETER BALL MILL

CUMULATIVE % PASSING

Size (Mesh)	5 Min	10 Min	20 Min	30 Min
65	68.68	92.73	99.93	100.00
100	60.83	89.42	99.86	100.00
150	52.03	81.90	99.18	99.96
200	42.54	70.29	96.25	99.42
325	33.32	56.58	87.26	94.89
400	30.38	51.96	82.60	91.63
Pan	0.00	0.00	0.00	0.00
% Solids	65	65	65	65

- Products of the five minute grind were analyzed to determine metal distrubutions.
- Products of the other grinds were not used in the metallurgical testwork.

PARTICLE DIZE DISTRIBUTIONS OF PLYMOUTH ORE GROUND IN 20 cm x 20 cm DIAMETER ROD MILL
CUMULATIVE % PASSING

Size (mesh)	P1 4 min	P1 (1) 4/3 min	P1 5 min	P-02 4/3 min	P-02 10 min	P-02 15 min	P-02 20 min
20		100.00					
28		99.64					
32	99.88						
35		98.06	100.00				
65	71.23	90.93	88.24	99.65			
100	36.90	71.11	68.02	92.28		95.44	96.15
150	78.83	51.94	53.60	78.43			
200	16.05	35.30	41.22	(66.00)(2)	85.45	86.31	91.85
270				51.30	68.11	81.37	83.47
325		29.54		46.03	58.59	73.07	80.15
400		23.06		42.45	52.08	65.74	69.81
Pan	0.00	0.00	0.00	0.00	0.00	0.00	0.00
% Solids	65	65	65	65	65	65	65

1. Reground + 65 mesh 3 minutes.
2. Obtained from % passing versus size plot used in ore flotation tests.
3. Ground product used in flotation tests MNF 01-32

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APPENDIX IV

*Gravity Testwork Mass Balances

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* Detailed mass balances for all spiral and flotation testwork are on file in Witteck's project records.

CALCULATION OF HEAVY LIQUID METAL RECOVERIES OF GROUND P-1 Ore

Size	Feed					Float Fraction		Sink Fraction							
	Wt	Assay %		Units (g)		Assay (%)		% Recov			Units (g)			Assays (%)	
(Mesh)	(g)	Mn	Fe	Mn	Fe	Mn	Fe	Wt	Mn	Fe	Wt	Mn	Fe	Mn	Fe
14	661.9	6.75	6.66	44.68	44.08	1.39	2.76	36.68	86.97	73.74	242.78	38.86	32.50	16.01	13.38
28	411.0	7.33	6.87	30.13	28.24	1.66	3.06	31.79	84.55	69.63	130.66	25.47	19.67	19.49	15.05
48	264.8	6.73	6.85	17.82	18.14	1.44	3.15	30.10	85.05	67.84	79.70	15.16	12.31	19.02	15.43
65	99.4	5.99	6.52	5.95	6.48	1.62	3.31	28.28	80.59	63.65	28.11	4.79	4.12	17.06	14.68
100	76.7	6.21	5.89	4.76	4.13	1.20	2.40	27.67	85.94	70.51	21.12	4.09	2.91	19.38	15.00
-100	465.0	5.38	5.94	25.02	27.62	5.33	5.88	1.13	2.11	2.07	5.25	0.53	0.57	10.03	10.90
TOTAL	1978.9	(6.49)	(6.51)	128.36	128.87	(2.65)	(3.86)	25.65	69.66	55.93	507.62	88.90	72.08	(17.51)	(14.20)

RESULTS OF VANNING TESTS OF P1 ORE

Test #	Van Type	Heavy Fraction (% Mn)
V1	25 cm diameter van	9.62
V2	mechanical vanner	6.78

Weights were not recorded

Feed was -6 mesh ore ground for five minutes in 500 gram ball mill at 65% solids.

RESULTS OF JIG TESTING OF COARSE FRACTION OF GROUND P1 ORE

Test		Wt	Assay (%)		Units (g)		% Dist'n		
#	Product	(g)	Mn	Fe	Mn	Fe	Wt	Mn	Fe
V1*	Bed	137	10.42	11.42	14.28	15.65	14.91	16.17	18.69
	Conc	22	13.63	25.90	3.00	5.70	2.40	3.40	6.81
	Tail	760	9.35	8.21	71.03	62.37	82.69	80.43	74.50
	TOTAL	919	9.61	9.11	88.31	83.72	100.00	100.00	100.00
V2*	Bed	73	7.02	6.44	5.12	4.70	11.57	10.95	9.79
	Conc	106	13.78	15.89	14.61	16.84	16.80	31.24	35.06
	Tail	452	5.98	5.86	27.03	26.49	71.63	57.81	55.15
	TOTAL	631	7.41	7.61	46.76	48.03	100.00	100.00	100.00

Recovery in Bed and Conc in Test 1 = 17.31% Wt, 19.57% Mn, 25.50% Fe
Conc. Grade in Bed and Conc in Test 1 = 10.87%Mn, 13.43% Fe

Recovery in Bed and Conc in Test 2 = 28.37% Wt, 42.19% Mn, 44.85% Fe
Conc. Grade in Bed and Conc in Test 2 = 11.02% Mn, 12.03% Fe

* - Jig feed was +65 mesh fraction of material ground for five minutes in 500 gram ball mill.

RESULTS OF TABLING OF GROUND P1 ORE

Test		Wt	Assay %		Units (g)		% Dist'n		
#	Product	(g)	Mn	Fe	Mn	Fe	Wt.	Mn	Fe
T1*	Concentrate	1968.5	12.08	9.25	237.79	182.09	21.37	35.72	28.75
	Middlings	3600.0	7.79	7.25	280.44	261.00	39.08	42.12	41.22
	Tailings	3643.1	4.05	5.22	147.55	190.17	39.55	22.16	30.03
	TOTAL	9211.6	(7.23)	6.87)	665.78	633.26	100.00	100.00	100.00
T2*	Concentrate	3661.4	8.24	8.18	301.70	299.50	33.59	44.41	41.92
	Middlings	3915.5	5.90	6.00	231.01	234.93	35.91	34.01	32.32
	Tailings	3324.4	4.41	5.54	146.61	184.17	30.50	21.58	25.77
	TOTAL	10901.3	(6.23)	(6.59)	679.32	718.60	100.00	100.00	100.00

Recovery in Conc + middling in Test 1 = 60.45% wt, 77.84 % Mn, 69.97% Fe
Product Grade of Conc + middling in Test 1 = 9.31% Mn, 7.03% Fe

Recovery in Conc + middling in Test 2 = 69.50% Wt, 78.42% Mn, 74.24% Fe
Product Grade of Conc + middling in Test 2 = 7.96% Mn, 7.00% Fe

* - Table feed was -6 mesh ore ground for 2 minutes in 2000 gram rod mill at 80% solids.

APPENDIX V

High Intensity Magnetic Separation Test Conditions and Mass Balances

WET HIGH INTENSITY MAGNETIC SEPARATION

(TEST WHIMS I)

Objective: To concentrate manganese minerals by wet intensity magnetic separation (WHIMS) with a coarse matrix.

Conditions:

- Box mag high intensity magnetic separator.
- Coarse matrix (6-8 mm approx.)
- Field strength 4-16 kilogauss
- Feed passed at 33 mL/second at 4 kilogauss
- The nonmagnetics from the first pass were repassed at 11 kilogauss at 33 mL/second
- The nonmagnetics from the second pass were repassed at 16 kilogauss at 33 mL/second.

Observations:

- Magnetics darker brown than nonmagnetics.

Conclusions:

- Magnetic product assayed 10.4 - 11.2% Mn.
- Mn recovery in magnetics was as high as 73.6% in 52.6% of the weight. Processing in this manner was therefore not superior to gravity or flotation processing.
- Fe recovery was very similar to Mn.

WET HIGH INTENSITY MAGNETIC SEPARATION

(TEST WHIMS II)

Objective: To concentrate manganese minerals by wet high intensity magnetic separation (whims) with a fine matrix.

Conditions:

- Sala high intensity magnetic separator at CANMET
- Field strength 4200-6800 gauss
- Fed - +200, -200, +400, -400 mesh fractions separately (overall 78% -270 mesh)
- Flow velocity was 105-144 mm/sec
- Used three matrices for the 3 products- #56 for +200, mesh, #8 for -200 mesh and 3/0 for -400 mesh

Observations:

- Magnetics darker brown than non-magnetics
- Decreased flowrate and increased field strength resulted in lower grade and higher Mn and weight recovery

Conclusions:

- Magnetics grade 10.3-14.8% with recoveries of 67-93% Mn in 32-66% of the weight.
- Overall recovery 81.4% in 46.1% of weight with product grade of 12.3%
- Results were better than those of initial (Whims I) testwork, but grades were not higher than those achieved with gravity/ flotation tests.
- Cleaning and finer grinding may be advantageous; however, further testwork was not planned.

APPENDIX VI

Mineralogy

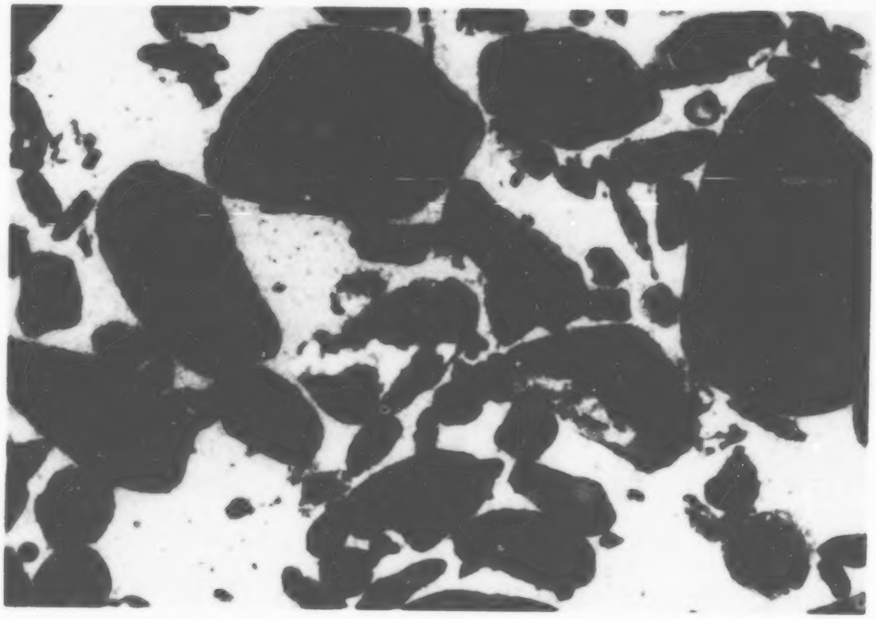

Photo 1 - 50x - Transmitted Light - 1cm = 150 microns
Spiral concentrate with abundant subrounded
oxide-rich particles and minor
carbonate-chlorite-silicate particles.

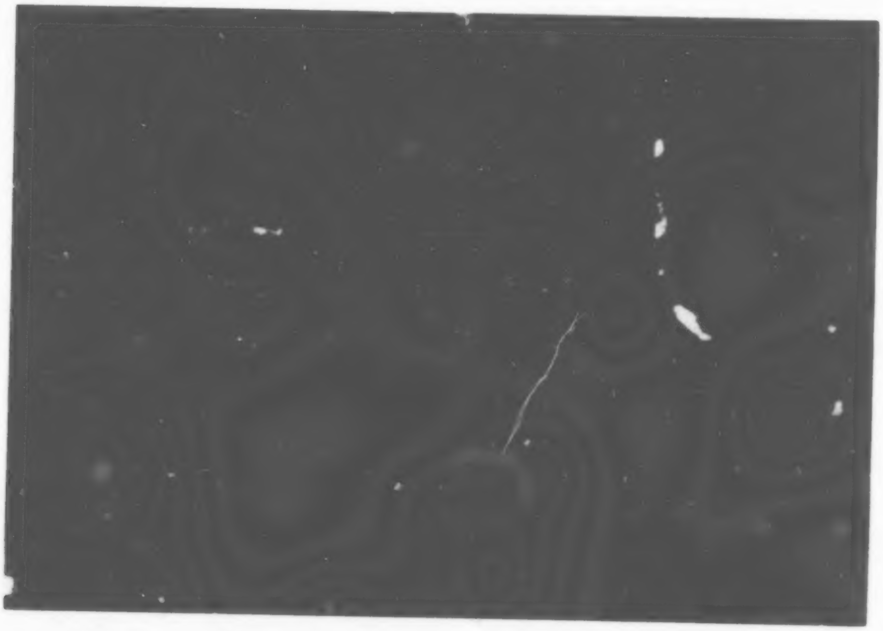

Photo 2 - 50x - Transmitted Light (Crossed Nicols)
1 cm - 150 microns
Spiral middlings product contains numerous subangular to
subrounded oxide-bearing particles, brown hematitic slate
chlorite-silicates, and minor liberated carbonates.

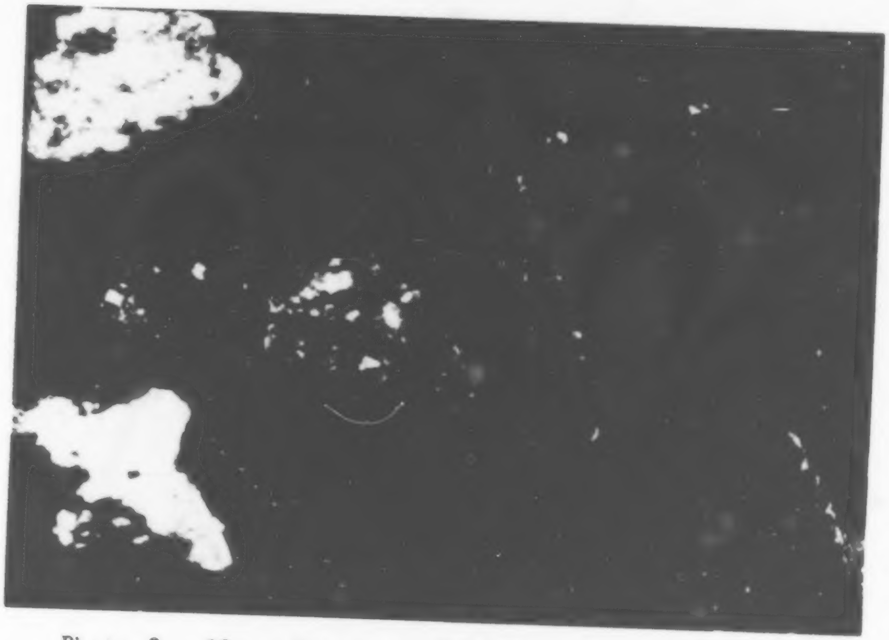

Photo 3 - 50x - Transmitted Light - (Crossed Nicols)
1 cm = 75 microns
Spiral tailings product containing numerous silicate and carbonate particles with fewer oxide-bearing particles.

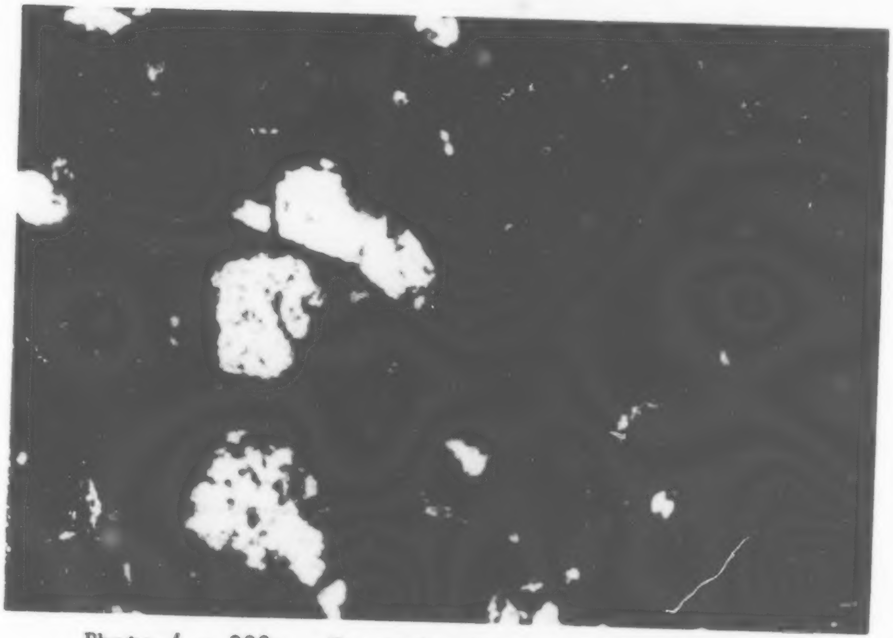

Photo 4 - 200x - Transmitted Light (Crossed Nicols)
1 cm - 40 microns
Reverse flotation tailings product containing abundant angular liberated fragments of carbonates and dark oxide-rich particles.

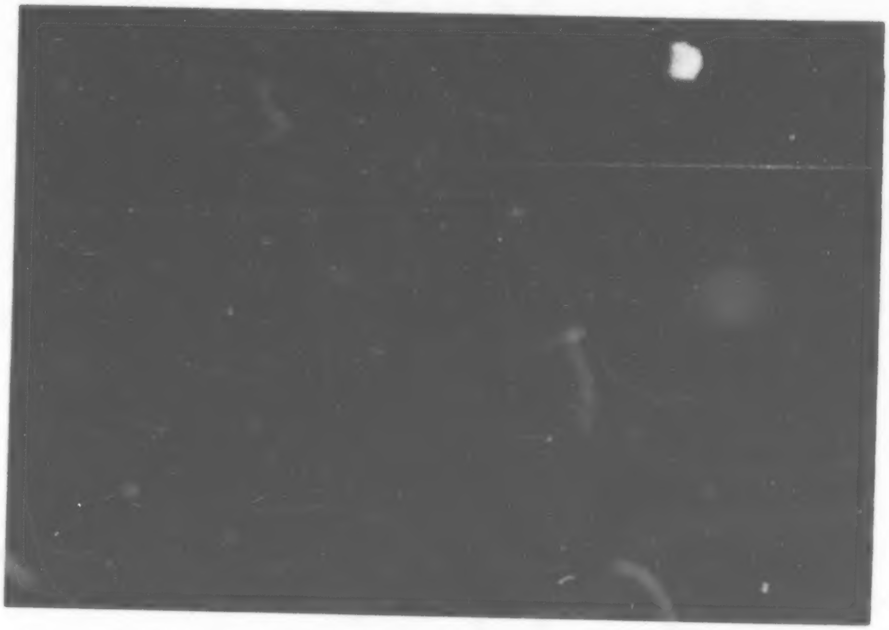

Photo 5 - 200x - Reflected Light (Crossed Nicols)
1 cm = 40 microns
Reverse flotation tailings product containing minor Mn oxide rich particles and abundant hematitic slates and other hematite-stained silicates.

Photo 6 - 100x - Transmitted Light (Crossed Nicols)
1 cm = 75 microns
Direct flotation concentrate product (subsequent to reverse flotation) exhibits abundant twinned carbonate particles with lesser dark oxide-rich particles.

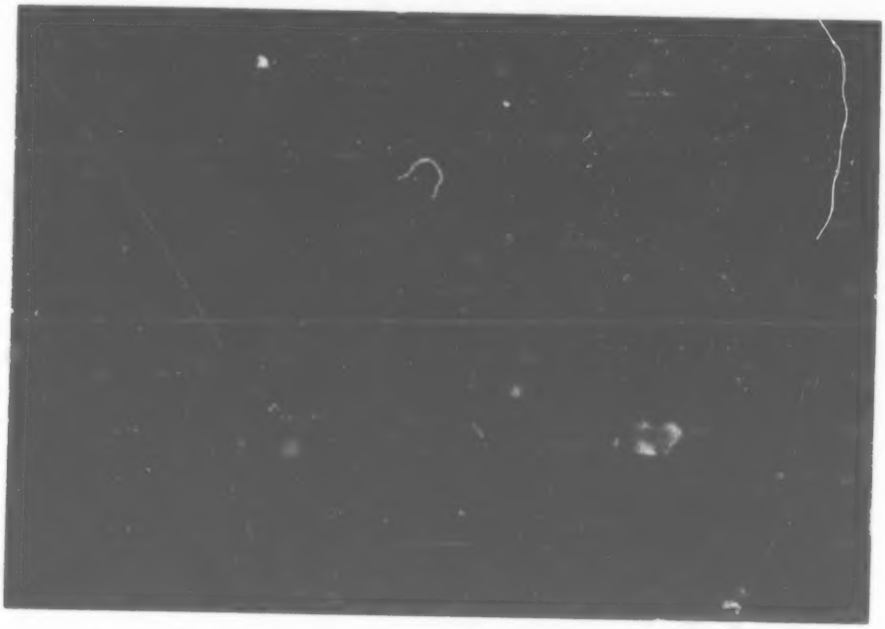

Photo 7 - 200x - Reflected Light (Crossed Nicols)
1 cm = 40 microns
Direct flotation concentrate No. 2 (subsequent to reverse flotation) exhibits abundant coarse oxide-rich particles, coarse liberated carbonates and minor hematitic slimes and silicates.

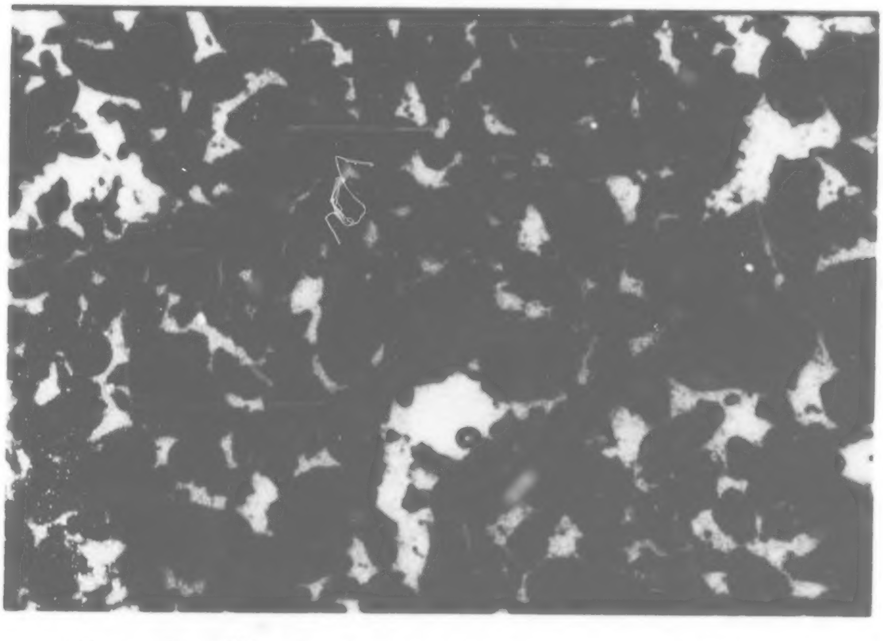

Photo 8 - 50x- Transmitted Light - 1 cm = 150 microns
Direct flotation tailings (subsequent to reverse flotation) exhibits numerous opaque particles and abundant chlorite-carbonate-silicate gangue particles.

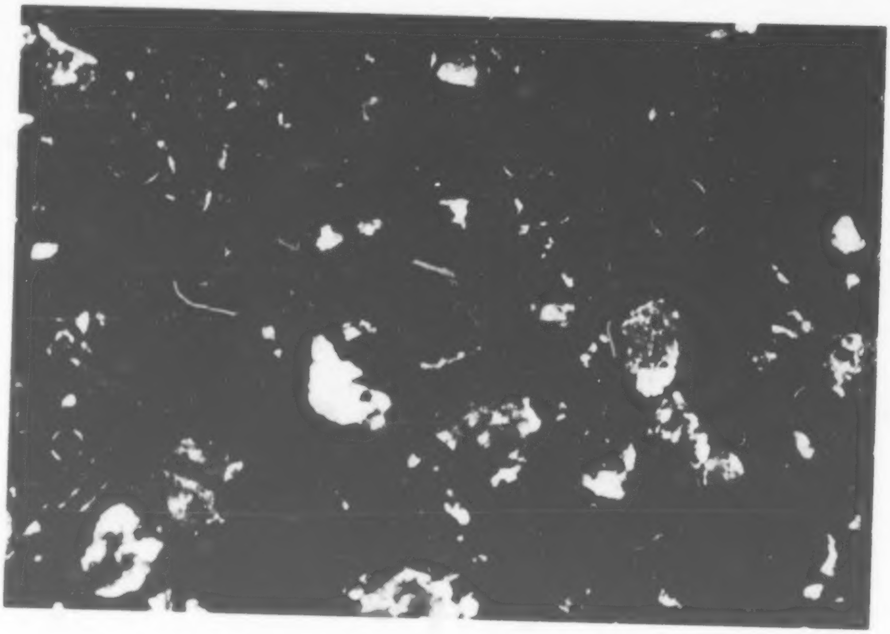

Photo 9 - 50x - Transmitted Light - (Crossed Nicols)
1 cm = 150 microns
Cleaner concentrate (direct flotation after reverse flotation) exhibits abundant liberated carbonate and oxide-rich particles.

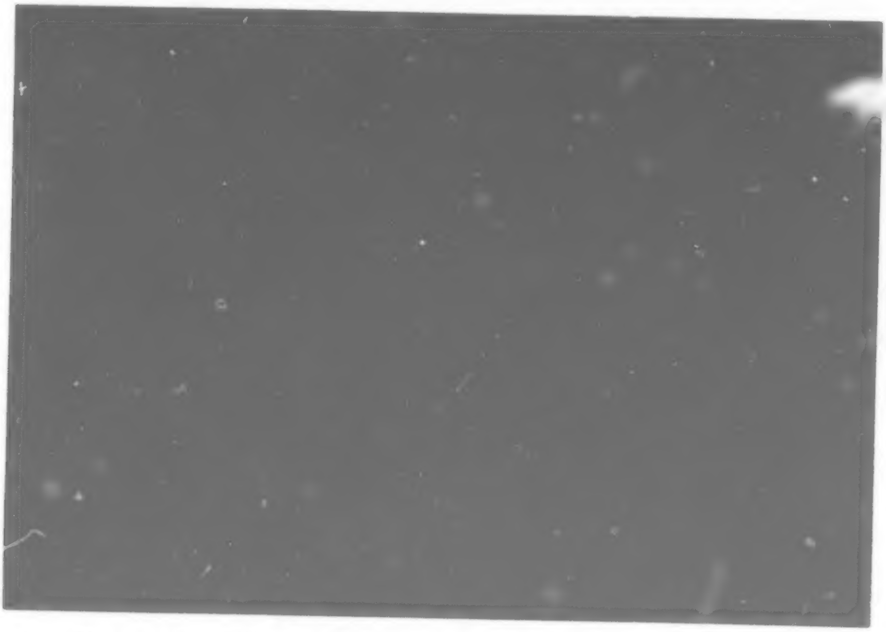

Photo 10 - 100x - Reflected Light - 1 cm = 75 microns
Cleaner tailings (direct flotation after reverse flotation) exhibits numerous Mn-oxide bearing particles with very high to poor degrees of liberation.
Dark coloured particles are primarily silicate gangue.

APPENDIX VII

Extractive Metallurgy Test Conditions and Mass Balances

FLOTATION PRODUCT SULPHATIZING ROAST

(Tests MEX A1 and A2)

Test Objectives:	To perform a sulphate roast on the flotation product composite, followed by manganese dissolution in water.
Equipment:	o Pyradia muffle furnace o Porcelain crucible o Leach pot and agitator
Conditions:	o Na_2SO_3 (sodium sulphite) was SO_2 source. o Addition of 240 kg/t of Na_2SO_3 to flotation product in Test MEX A1. o Addition of 504 kg/t of Na_2SO_3 to flotation product in Test MEX A2. o Roast at 400° C for 2 hours, then at 675°C for 3.5 hours. o Leach at 1% solids for 20 hours in water. o Pregnant solution assayed for Mn at 2 and 20 hours.
Conclusions:	o Virtually none of the manganese dissolved. o No further tests were planned.

FLOTATION PRODUCT OXIDIZING ROAST

(Test MEX B)

Test Objectives:	To perform an oxidizing roast on the flotation product composite, followed by manganese dissolution in water.
Equipment:	o Pyradia muffle furnace o Porcelain crucible o Leach pot and agitator
Conditions:	o Addition of 384 kg/t of H_2SO_4 and 480 kg/t of NaCl to flotation product. o Roast at 500°C for 1 hour. o Leach at 1% solids for 20 hours in water. o Pregnant solution assayed for Mn at 3 and 20 hours.
Conclusions:	o Mn extraction at 3 hours was 22.9% with 314 ppm Mn in solution. o Mn extraction at 20 hours was 23.7%, with 326 ppm in solution.

FLOTATION PRODUCT REDUCING LEACH

(Test MEX C)

Test Objectives: To perform a reducing leach with SO_2 and H_2SO_4 on the flotation product composite.

Equipment:

- Leach pot and agitator
- SO_2 cylinder and glass frit

Conditions:

- Leachant prepared by adjusting pH to 3 with H_2SO_4; SO_2 was bubbled in to maintain a pH of no less than 3; if the pH fell below 3, gas addition was stopped.
- The composite was added; pulp density was 1% solids; SO_2 addition continued to maintain pH 3.
- Pregnant solution assayed at 2, 4 and 6 hours.

Conclusions:

- Mn extraction was 45.6%, 63.0% and 71.7% after 2, 4 and 6 hours.
- Mn tenor in solution was 501, 693, and 789 ppm after 2, 4 and 6 hours.
- Mn extraction was high; a longer retention time may have resulted in increased extraction.
- These data were useful in providing a starting point for subsequent treatment of the Plymouth ore.

COLUMN HEAP LEACH
(TEST MHL 1)

Test Objectives: To heap leach 0.95 cm ore (Plymouth 1-4 composite) with sulphuric acid and sulphur dioxide.

Equipment:

- 5.0 cm diameter plastic column
- SO_2 cylinder and flowmeter
- Peristaltic pump, agitator

Conditions:

- Agglomerated with sulphuric acid and cured 48 hours.
- Solution changed after 10 days and partially changed after 17 days
- SO_2 bubbled into tank; agitator aided in gas adsorption.

Observations:

- Mn dissolution slow at pH3 and low SO_2 rate
- Initial solution change may have aided extraction but pH was lowered below 2 concurrently.
- Percolation rate was quite variable as the column would plug and unplug. Rates were, however, quite high at 0.37-1.17 USG/ft^2/min.

Conclusions:

- Mn extraction was 71% after 49 days; leaching was essentially complete after 33 days.
- Mn tenor in solution was 3 gpl
- Iron extraction was 19.5% ; iron tenor in solution was 1.5 gpl; Fe tenor in solution at pH3 was only 0.1 gpl.
- SO_2 addition was inefficient and is not indicative of actual heap reagent consumptions.

MASS BALANCE OF HEAP LEACH MHL1

Product	Wt/Vol kg/L	Assay, gpl %		Units, g		% Dist'n	
		Mn	Fe	Mn	Fe	Mn	Fe
Filtrate 1	23.0 (1)	3.48	0.10	80.04	2.30	27.69	0.66
Lost Sol'n	17.6 (2)	3.03	1.58	53.33	27.81	18.45	8.01
Filtrate 2	23.0 (3)	3.09	1.58	71.07	36.34	24.58	10.46
Wash 1	5.9 (4)	0.35	0.19	2.07	1.12	0.72	0.32
Wash 2,3	7.1	0.06	-	0.43	-	0.15	-
Residue	3.02	2.72	7.61	82.14	279.82	28.41	80.55
Total		(7.32)	(8.80)	289.08	347.39	100.00	100.00

	Mn	Fe	Wt
Assayed Head	7.57%	9.46%	
Calculated Head	7.32%	8.80%	
Mass Balance	96.70%	93.02%	
Weight Loss			32.5%
Extraction:	71.59%	19.45%	

1) 10 days,
2) 17 days,
3) 49 days,
4) Wash 1, 2 = pH 2.5 H_2SO_4; Wash 3 = Water.

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METAL DISTRIBUTIONS IN THE MHL 1 RESIDUE

Size			% Distribution				
	Assay (%)		In Residue			From Ore	
(mesh)	Mn	Fe	Wt.	Mn	Fe	Mn	Fe
3	3.03	6.84	26.89	33.53	26.40	9.54	21.27
4	2.82	7.07	20.99	24.37	21.29	6.92	17.14
8	2.52	6.98	24.83	25.73	24.85	7.30	20.02
-8	1.46	7.02	27.29	16.37	27.46	4.65	22.12
Total:	(2.43)	(6.97)	100.00	100.00	100.00	28.41	80.55

Residue Assayed Head = 2.72 % Mn, 8.80 % Fe

Mass Balance = 89.34 % Mn, 79.20 % Fe

BOTTLE LEACH TESTS

(TESTS MBL 1-4)

Objective: To investigate the effect of pH and reagent levels in simulated heap leach tests.

Conditions:
- o Feed 0.95 cm ore (as for heap leach)
- o Leach in sealed 2.2 L Winchester bottles at 28 rpm
- o Staged additions sodium metabisulphite ($Na_2S_2O_5$) as sulfur dioxide source
- o In two tests, solution was changed after 6 days of leaching.
- o Total retention time, 5-14 days
- o Test conditions were as follows:

Test #	pH	Total H SO Add'n (kg/t)	Total SO_2 Addn'n (% of Stoich)	Individual SO_2 Add'n (% of Stoich)
MBL-1	1.5	127.8	-	-
2	2.5	149.2	345	35
3	2.5	162.9	862	86
4	1.5	212.0	431	86

Observations: Filtration rates were very slow; autogenous grinding of the ore occurred as the test proceeded.

Conclusions:

- There was an upper limit to practical reagent additions (i.e. pH 1.5 and 86% of stoichiometric per addition resulted in uncontrollable fizzing and loss of material).
- Sulphuric acid alone gave only 27% Mn extraction, presumably mainly due to carbonate leaching.
- SO_2 additions greater than 345% of stoichiometric did not result in increased Mn dissolution.
- Maximum manganese extraction was 59% in 14 days, which is comparable to heap leach results.

MASS BALANCE OF HEAP LEACH MHL1

Product	Wt/Vol kg/L	Assay, gpl %		Units, g		% Dist'n	
		Mn	Fe	Mn	Fe	Mn	Fe
Filtrate 1	23.0 (1)	3.48	0.10	80.04	2.30	27.69	0.66
Lost Sol'n	17.6 (2)	3.03	1.58	53.33	27.81	18.45	8.01
Filtrate 2	23.0 (3)	3.09	1.58	71.07	36.34	24.58	10.46
Wash 1	5.9 (4)	0.35	0.19	2.07	1.12	0.72	0.32
Wash 2,3	7.1	0.06	-	0.43	-	0.15	-
Residue	3.02	2.72	7.61	82.14	279.82	28.41	80.55
Total		(7.32)	(8.80)	289.08	347.39	100.00	100.00

	Mn	Fe	Wt
Assayed Head	7.57%	9.46%	
Calculated Head	7.32%	8.80%	
Mass Balance	96.70%	93.02%	
Weight Loss			32.5%
Extraction:	71.59%	19.45%	

1) 10 days,
2) 17 days,
3) 49 days,
4) Wash 1, 2 = pH 2.5 H_2SO_4; Wash 3 = Water.

SERIES A
BOTTLE LEACH TEST
MASS BALANCE

Test MBL 1

Product	L/grams	% gpl		Units, g		% Dist.	
		Mn	Fe	Mn	Fe	Mn	Fe
Filtrate 1	800	15.92		12.74		19.18	
Wash 1	1028	4.47	1.08	4.60		6.92	
Filtrate 2	-	-	-	-	-	-	
Wash 2	1010	0.47	0.08	0.47		0.71	
Wash 3	580	0.16	-	0.09		0.14	
Residue	876.1	5.54		48.54		73.05	
Total		(6.67)		66.44		100.00	

Feed: 996 g
Assayed Head: 7.57% Mn 9.46% Fe
Calculated Head: 6.67% Mn
Mass Balance: 88.11% Mn
% Extraction: 26.95% Mn

Weight Loss: 12.0%

SERIES A
BOTTLE LEACH TEST
MASS BALANCE

Test MBL 2

		% gpl		Units, g		% Dist.	
Product	L/grams	Mn	Fe	Mn	Fe	Mn	Fe
Filtrate 1	860	16.98		14.60		19.06	
Wash 1	3540	2.62		9.27		12.10	
Filtrate 2	976	15.65	4.42	15.27		19.93	
Wash 2	997	5.09	1.48	6.07		6.62	
Wash 3	1001	1.09	0.19	1.09		1.42	
Residue	824	3.80	7.15	31.31		40.87	
Total		(2.63)		76.61		100.00	

Feed: 1004 g

Assayed Head: 7.57% Mn 9.46% Fe

Calculated Head: 7.63% Mn

Mass Balance: 100.79% Mn

% Extraction: 59.13% Mn

Weight Loss: 17.9%

SERIES A
BOTTLE LEACH TEST
MASS BALANCE

Test MBL 3

Product	L/grams	% gpl		Units, g		% Dist.	
		Mn	Fe	Mn	Fe	Mn	Fe
Filtrate 1	865	11.91		10.30		15.95	
Wash 1	1580	1.24		1.96		3.04	
Filtrate 2	1004	16.06		16.12		24.97	
Wash 2	1026	5.78		5.93		9.18	
Wash 3	1019	2.46		2.51		3.89	
Residue	802	3.46		27.75		42.97	
Total		(6.41)		64.57		100.00	

Feed: 1007 g

Assayed Head: 7.57% Mn 9.46% Fe

Calculated Head: 6.41% Mn

Mass Balance: 84.67% Mn

% Extraction 57.03% Mn

Weight Loss: 20.4% ~~Mn~~

SERIES A
BOTTLE LEACH TEST
MASS BALANCE

Test MBL 4

Product	L/grams	% gpl		Units, g		% Dist.	
		Mn	Fe	Mn	Fe	Mn	Fe
Filtrate 1	1000	31.92	-	31.92	-	42.48 *	-
Wash 1							
Filtrate 2							
Wash 2							
Wash 3							
Residue							
Total							

Feed:

Assayed Head: % Mn % Fe

Calculated Head: % Mn % Fe

Mass Balance: % Mn % Fe

% Extraction 42.48 Mn

Weight Loss: NA

* Based upon assayed head; no other volumes, weights assays taken.

BOTTLE LEACH TESTS

(TESTS MBL 5-8)

Objective: To isolate the effects of major leaching parameters using consistent testing procedures. The following variables were investigated:

- Particle size - ground versus crushed
- pH, 2.0 or 3.0
- Effect of change of solution (i.e. Mn tenor)
- Effect of lower SO_2 levels

Conditions:

- Leach in 2.2 L Winchester bottles at 28 rpm
- Solution changed approximately each 48 hours (if applicable), followed by reagent addition
- Total retention time 10 days
- Ground ore, 85% -200 mesh (for 1 test)
- Crushed ore, - 0.95 cm

Observations:

- Filtration rates were very low; the ground ore filtered the fastest at the end (slowest at start); this was due to autogenous grinding of the crushed ore.

Conclusions:

- Maximum Mn extraction was obtained for the ground ore (80%); maxiumum for the crushed ore was 57% under the same conditions.
- Those values were achieved with 100% stoichiometric SO_2.
- Changing of solution resulted in a 5% increase in Mn extraction (57% versus 52%); the difference in leachability started at 22 gpl Mn.
- Leaching at pH 2 versus pH3 resulted in an increase in Mn extraction from 39 to 57%; therefore H_2SO_4 mixture is definitely required to assist the sulphurous acid lixiviant.
- Dissolution of Fe was only 2-3% in two tests.

SERIES B
BOTTLE LEACH TEST
MASS BALANCE

Test MBL 5

Product	L/grams	% gpl		Units, g		% Dist.	
		Mn	Fe	Mn	Fe	Mn	Fe
Filtrate 1	0.010	9.37		0.09		0.22	
Filtrate 2	0.320	27.39		8.76		21.37	
Filtrate 3	0.365	23.14		8.45		20.61	
Filtrate 4	0.380	15.74		5.98		14.59	
Filtrate 5	0.390	13.05		5.09		12.42	
Filrate 6	0.430	6.99		3.01		7.34	
Wash 1	1.000	1.62		1.62		3.95	
Wash 2							
Wash 3							
Residue	425	1.88		7.99	10.69	19.50	3.95
Total		(8.20)		40.99	100.00		

Feed: 500 grams
Assayed Head: 7.57 Mn 9.46% Fe
Calculated Head: 8.20% Mn - % Fe
Mass Balance: 108.3 % Mn - % Fe
% Extraction 80.50% Mn, 3.95% Fe

Weight Loss: 15.0%

SERIES B
BOTTLE LEACH TEST
MASS BALANCE

Test MBL 6

Product	L/grams	% gpl		Units, g		% Dist.	
		Mn	Fe	Mn	Fe	Mn	Fe
Filtrate 1	0.010	0.89		0.01		tr	
Filtrate 2	0.440	11.50		5.06		12.82	
Filtrate 3	0.470	11.98		5.63		14.28	
Filtrate 4	0.384	9.92		3.81		9.66	
Filtrate 5	0.420	10.80		4.54		11.51	
Filtrate 6	0.440	5.48		2.41		6.11	
Wash 1	1.000	1.19		1.19		3.02	
Wash 2							
Wash 3							
Residue	431	3.90		16.81		42.60	
Total		(7.89)		39.46		100.00	

Feed: 500 grams
Assayed Head: 7.57% Mn 9.46 % Fe
Calculated Head:7.89% Mn - % Fe
Mass Balance: 104.2% Mn - % Fe
% Extraction 57.40% Mn, - % Fe

Weight Loss: 13.8%

SERIES B
BOTTLE LEACH TEST
MASS BALANCE

Test MBL 7

		% gpl		Units, g		% Dist.	
Product	L/grams	Mn	Fe	Mn	Fe	Mn	Fe
Filtrate 1	0.010	0.27		0.00		tr	
Filtrate 2	0.430	8.96		3.85		10.71	
Filtrate 3	0.445	7.50		3.34		9.29	
Filtrate 4	0.427	4.69		2.00		5.56	
Filtrate 5	0.407	5.79		2.36		6.56	
Filtrate 6	0.525	2.68		1.41		3.92	
Wash 1	1.000	1.11		1.11		3.09	
Wash 2							
Wash 3							
Residue	459	4.77		21.89		60.87	
Total		(7.19)		35.96		100.00	

Feed: 500 grams

Assayed Head: 7.57% Mn 9.46 % Fe

Calculated Head: 7.19% Mn - % Fe

Mass Balance: 95.0% Mn - % Fe

% Extraction 39.13% Mn, - % Fe

Weight Loss: 8.2%

SERIES B
BOTTLE LEACH TEST
MASS BALANCE

Test MBL 8

Product	L/grams	% gpl		Units, g		% Dist.	
		Mn	Fe	Mn	Fe	Mn	Fe
Sample 1	0.010	0.89		0.01		0.03	
Sample 2	0.010	12.15		0.12		0.27	
Sample 3	0.020	21.65		0.43		1.14	
Sample 4	0.015	27.16		0.41		1.09	
Sample 5	0.020	33.60		0.67		1.78	
Filtrate 1	0.410	32.48	4.28	13.32	2.14*	35.30	4.32*
Wash 1	1.000	4.72		4.72		12.51	
Residue	446	4.05	10.62	18.06	47.37	47.86	95.68
Total		(7.55)	(9.90)*	37.74	49.51	100.00	100.00

Feed: 500 grams
Assayed Head: 7.57% Mn 9.46% Fe
Calculated Head: 7.55% Mn 9.90% Fe *
Mass Balance: 99.7 % Mn 104.65% Fe *
% Extraction 52.14% Mn, 4.32% Fe

Weight Loss: 10.8%

* Assuming final solution 500 mL liquid

Volume

2

5-9217

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PROCESS EVALUATION OF SELECTED OPTIONS FOR TREATING NEW BRUNSWICK MANGANESE MINERAL DEPOSITS:

VOLUME II
ECONOMIC ANALYSIS

Project 5211

for

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Place du Portage,
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Hull, Quebec
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1.0 SOMMAIRE

On a porté de cinq à dix le nombre d'options de processus de traitement du minerai de manganèse soumises à l'analyse économique, pour pouvoir effectuer une évaluation plus globale du potentiel du projet.

On a résumé à la table 1, "Investissement, frais d'exploitation et analyse", les options de processus de traitement étudiées et les résultats de l'analyse économique.

On a évalué les coûts d'investissement à partir des coûts des équipements en employant des facteurs multiplicatifs applicables aux traitements de lixiviation.

Les frais d'exploitation sont basées sur les prix courants des réactifs, sur les consommations de réactifs mesurées et calculées, et sur les consommations effectives de réactifs dans des projets similaires.

Les processus de traitement étudiés sont tous basés sur la lixiviation du minerai ou du concentré à l'aide d'acide sulfurique et de gaz sulfureux, avec maintien d'un milieu de lixiviation réducteur.

Pour pouvoir entreprendre l'évaluation économique, on devait d'abord choisir un produit hydrométallurgique de base et un processus de traitement utilisable en production. Comme produit de base, on a choisi un précipité chimique de qualité moyenne (pureté d'environ 95%), et le schéma du processus de traitement incorporait la technologie hydrométallurgique standard. Le processus comprenait:

- Broyage primaire du minerai
- Broyage semi autogène et broyage par broyeur à galets
- Épaisissement/sédimentation en lieu neutre et filtration
- Lixiviation conventionnelle
- Filtration du résidu de lixiviation acide
- Neutralisation et rejet des résidus du traitement

On a utilisé un processus de purification de la solution de $CaCl_2$ pour éliminer sulfate et gypse avant la précipitation du manganèse sous forme de $Mn(OH)_2$ avec de la chaux. Le réactif chloré a été concentré par évaporation avant recyclage.

Toutes les options de processus étudiées dérivaient du cas de base ci-dessus, par substitution de nouvelles options de circuit et élimination de circuits jugés inutiles. On ajustait les circuits conservés pour tenir compte des modifications du minerai et des débits de solution.

Dans le volume 1 de ce rapport, on a montré que le traitement de valorisation du minerai de manganèse ou sa pré-concentration pour l'obtention d'une matière première utilisable par le traitement pyrométallurgique ultérieur ne donnait pas un résultat satisfaisant. Cependant, la pré-concentration permettait de réduire la quantité de gangue (consommatrice d'acide) et la consommation de réactifs au cours de la lixiviation. On a donc conservé cette option potentiellement utile pour la recherche d'un processus de traitement économique.

L'option de pré-concentration comprenait des concentrateurs à spirales entre le broyage semi autogène et les broyeurs à galets. Dans le cas de l'option de lixiviation en tas, la lixiviation du matériau empilé remplaçait les opérations d'épaississement/sédimentation en milieu neutre, lixiviation conventionnelle et filtration en milieu acide. L'option de récupération du métal par dépot électrolytique a permis l'emploi de cellules électrolytiques à la place du circuit de précipitation du produit utilisé dans le processus de base.

On a également modifié le processus de base pour inclure une combinaison de pré-concentration et lixiviation en tas. On a également évalué cette combinaison en conjonction avec les options de récupération par dépôt électrolytique, extraction par solvant et production de sulfate de manganèse.

L'option "sulfate de manganèse" comprenait le recouvrement du produit par cristallisation à haute température, à la place de la précipitation. On a également évalué une unité de production "minimum" pour le recouvrement du sulfate de manganèse.

La nouvelle technologie proposée est basée sur purification et extraction par solvant puis recouvrement à partir de la solution, et elle tient compte de la précipitation du produit et de la récupération du réactif.

L'analyse économique indique que la production de sulfate de manganèse ($MnSO_4$) n'est pas rentable, car les frais d'exploitation sont supérieurs à la valeur du produit.

La production d'un composé chimique de qualité moyenne est également non rentable, car le total des frais de production est supérieur à la valeur estimée du produit. Il est possible que la valeur estimée du produit choisie pour cette étude soit supérieure à la valeur réelle.

La production de manganèse métallique de haute pureté (produit de grande valeur) par recouvrement électrolytique est marginalement rentable, mais seulement lorsqu'on inclut des opérations de pré-concentration et lixiviation en réacteur dans le processus de traitement. Cependant, on considère que la petite marge observée est insuffisante pour justifier le risque et la complexité de l'exploitation d'un circuit de récupération électrolytique du manganèse.

La meilleure rentabilité d'exploitation est fournie par un circuit hydrométallurgique qui comprend pré-concentration et lixiviation en réacteur, et une nouvelle technologie d'extraction par solvant pour la purification et le recouvrement d'un produit de haute pureté. La validité technique d'une telle approche a été mise en évidence dans d'autres projets relativement similaires. La voie de l'extraction par solvant fournissait une marge d'exploitation positive lorsqu'on l'employait seule dans le processus de traitement hydrométallurgique standard.

Les coûts d'investissement étaient généralement élevés du fait de la nature acide des solutions utilisées, comme c'est généralement le cas dans de tels types de processus.

Dans la plupart des cas, les coûts d'exploitation élevés sont imputables aux coûts élevés et à la forte consommation des réactifs de lixiviation. La consommation est imposée par la minéralogie du minerai. Il serait éventuellement possible de réduire les coûts par la production locale d'acide sulfurique et gaz sulfureux.

L'expansion du domaine d'étude a permis de montrer que la combinaison de la pré-concentration et de la lixiviation en réacteur permet de conserver les coûts d'exploitation et d'investissement modérés inhérents à ces deux options. De même, on a pu voir que l'emploi de la nouvelle technologie basée sur l'extraction par solvant produisait la marge d'exploitation la plus encourageante.

Le concept sous-jacent au processus recommandé est qu'il est possible de produire un précipité chimique dont la pureté (99% minimum) et la valeur marchande sont similaires à celle du manganèse électrolytique, tandis que les coûts de production et les risques du processus sont inférieurs. La production d'un composé sulfurique comme le sulfate cristallisé limiterait le marché; on ne devrait pas utiliser cette option. Cependant, la production d'un carbonate ou d'un hydroxyde pourrait être facilement adaptée à plusieurs marchés comme:

- marché des piles au MnO_2
- marché des alliages métalliques de haute pureté

Table I

INVESTISSEMENT, FRAIS D'EXPLOITATION ET ANALYSE

Options de processus de traitement	Coût d'investissement (milliers de $)	Coût d'investissement Amort. et intérêts $/kg Mn	Frais d'exploitation milliers de $ par année	Frais d'exploitation $/kg Mn	Frais d'extr. minière $/kg Mn	Coût total du produit	Valeur du produit $/kg Mn	Marge d'exploitation $/kg Mn
1. Traitement hydrométallurgique standard, pour la production de $Mn(OH)_2$ de qualité moyenne	33,871	1,02	18,118	1,81	0,05	2,88	2,00	-0,88
2. Effets de la pré-concentration sur le traitement hydrométallurgique standard	29,232	0,88	12,946	1,30	0,10	2,28	2,00	-0,28
3. Effets de la lixiviation en tas sur le traitement hydrométallurgique standard	22,116	0,66	19,184	1,92	0,06	2,64	2,00	-0,64
4. Effets de la récupération électrolytique sur le traitement hydrométallurgique standard	52,871	1,59	23,823	2,38	0,05	4,02	3,85	-0,17
5. Pré-concentration et lixiviation en réacteur pour le traitement hydrométallurgique standard, production de $Mn(OH)_2$	22,823	0,69	13,960	1,40	0,12	2,21	2,00	-0,21
6. Pré-concentration et lixiviation en réacteur pour le traitement hydrométallurgique standard, avec récupération électrolytique	44,043	1,32	18,714	1,87	0,12	3,31	3,85	+0,54
7. Pré-concentration et lixiviation en réacteur pour le traitement hydrométallurgique standard, avec extraction par solvant pour production d'un produit de haute pureté	23,915	0,72	16,712	1,67	0,12	2,51	3,85	+1,34
8. Traitement hydrométallurgique conventionnel pour $MnSO_4$	33,352	1,00	15,585	1,56	0,05	2,61	1,06	-1,55
9. Traitement hydrométallurgique simplifié pour $MnSO_4$	25,798	0,77	14,991	1,50	0,05	2,32	1,06	-1,26
10. Pré-concentration et lixiviation en réacteur pour le traitement hydrométallurgique conventionnel, pour $MnSO_4$	22,104	0,66	11,637	1,16	0,12	1,94	1,06	-0,88

1.0 SUMMARY

The number of process alternatives for manganese ore processing to be analyzed economically was expanded from five to ten to permit a more comprehensive evaluation of the project potential.

The process alternatives studied and results of the economic analysis are summarized in Table 1, "Capital Cost, Operating Cost and Analysis".

Capital costs were factored from equipment costs using factors applicable to leaching processes.

Operating costs were based on current reagent prices, measured and calculated reagent consumptions, and operating requirements for similar projects.

The processes studied were all based on leaching the ore or concentrate using sulphuric acid and sulphur dioxide to maintain reducing leach conditions.

Selection of a base case hydrometallurgical product and a production process flowsheet were required to start the economic evaluation. A middling grade (about 95% purity) chemical precipitate was chosen as the product and the production flowsheet incorporated standard hydrometallurgical process technology. This flowsheet included:

- Primary ore crushing
- SAG and ball mill grinding
- Neutral thickening and filtration
- Conventional leaching
- Acid leach residue filtration
- Tailings neutralization and disposal.

A $CaCl_2$ solution purification process was used to remove sulphate as gypsum prior to manganese precipitation as $Mn(OH)_2$ with lime. Chloride reagent was concentrated by evaporation before being recycled to the process.

All the alternative processes studied were derived from the above base case, with new circuit options substituted into the flowsheet, and unnecessary circuits deleted. Remaining circuits were then adjusted for changes to ore and solution flow rates.

In Volume I of this report, it was shown that mineral processing or pre-concentration of manganese ore to achieve feed grades for subsequent pyrometallurgical treatment was not successful. However, pre-concentration did reduce acid consuming gangue and the consumption of reagents during leaching. Therefore it was retained as a potentially useful option in the search for an economically attractive process flowsheet.

The pre-concentration alternative included spiral concentrators between the SAG and ball mills. In the heap leaching option, the heaps replaced the neutral thickening, conventional leaching and acid filtration circuits. The electrowinning option provided for metal recovery in electrolytic cells to replace the product precipitation circuit of the base case flowsheet.

The base case flowsheet was also modified to include a combination of pre-concentration and heap leaching. This combination was also evaluated in conjunction with the electrowinning, solvent extraction and the manganese sulphate options.

The manganese sulphate options included product recovery by high temperature crystallization in place of precipitation. A "minimal" plant for manganese sulphate recovery was also evaluated.

The postulated new technology was based on solvent extraction purification and recovery from solution with allowances for product precipitation and reagent recovery.

The economic analysis indicated that production of manganese sulphate ($MnSO_4$) is uneconomic, as operating costs are greater than the product value.

Production of a middling grade chemical product was also uneconomic, as total production costs were greater than the assumed product value. The product value in this study may have been overstated compared to the real value.

Electrowinning to produce high purity manganese metal (a high value product) was marginally economic, but only then when the pre-concentration and vat leaching options were included in the flowsheet. However, the small operating margin was considered insufficient to compensate for the risk and complexity of operating a manganese electrowinning circuit.

A hydrometallurgical circuit which included pre-concentration, and vat leaching, and new solvent extraction technology for purification and recovery of a high purity product provided the best operating margin. The technical validity of such an approach has been demonstrated for other generally similar projects. The solvent extraction route still provided a positive operating margin when included singly in the standard hydrometallurgical flowsheet.

Capital costs were generally high due to the acidic nature of the solutions handled, but this is typical for these types of processes.

The high operating costs in most cases are a function of high leach reagent consumption and costs. Consumption is dictated by the minerology of the ore. Some reductions in costs might be possible by local manufacturing of sulphuric acid and sulphur dioxide.

Expansion of the scope of work showed that combining pre-concentration and vat leaching preserved the low operating and capital cost features inherent in the two alternatives. Also, it was shown that the new technology route based on solvent extraction provided the most encouraging operating margin.

The concept behind the recommended process route is that a chemical precipitate can be produced which has the purity (99% minimum) and market value of electolytic manganese but at lower production cost and operating risk. Sulphur bearing products such as sulphate crystal would limit market scope and should not be persued. However, a carbonte or hydroxide product form may be readily adapted to several markets, such as:

- High-tech MnO_2 battery market
- High purity metal alloy market

Table I

CAPITAL COST, OPERATING COST AND ANALYSIS

Process Flowsheet Alternative	Capital Cost		Operating cost		Mining Cost $/kg Mn	Total Product Cost	Product Value $/kg Mn	Operating Margin $/kg Mn
	$ (000's)	Deprec. & Interest $/kg Mn	$/year (000's)	$/kg Mn				
1. Standard Hydrometallurgical Process for Production of a Middling Chemical Product $Mn(OH)_2$	33.871	1.02	18.118	1.81	0.05	2.88	2.00	-0.88
2. Effects of Pre-concentration on the Standard Hydrometallurgical Process	29.232	0.88	12.946	1.30	0.10	2.28	2.00	-0.28
3. Effects of Heap Leaching on the Standard Hydrometallurgical Process	22.116	0.66	19.184	1.92	0.06	2.64	2.00	-0.64
4. Effects of Electrowinning on the Standard Hydrometallurgical Process	52.871	1.59	23.823	2.38	0.05	4.02	3.85	-0.17
5. Pre-concentration and Vat Leaching for the Standard Hydrometallurgical Process for $Mn(OH)_2$	22.823	0.69	13.960	1.40	0.12	2.21	2.00	-0.21
6. Pre-concentration and Vat Leaching for the Standard Hydrometallurgical Process with Electrowinning	44.043	1.32	18.714	1.87	0.12	3.31	3.85	+0.54
7. Pre-concentration and Vat Leaching for the Standard Hydrometallurgical Process with Solvent Extraction for Production of a High Purity Product	23.915	0.72	16.712	1.67	0.12	2.51	3.85	+1.34
8. Conventional Hydrometallurgical Process for $MnSO_4$	33.352	1.00	15.585	1.56	0.05	2.61	1.06	-1.55
9. Stripped Hydrometallurgical Process for $MnSO_4$	25.798	0.77	14.991	1.50	0.05	2.32	1.06	-1.26
10. Pre-concentration and Vat Leaching for the Conventional Hydrometallurgical Process for $MnSO_4$	22.104	0.66	11.637	1.16	0.12	1.94	1.06	-0.88

2.0 INTRODUCTION

The initial work performed on the New Brunswick Plymouth manganese ore under contract to the Department of Supply and Services, while well executed, did not yield the expected technical results. A concentrate could not be produced using mineral processing methods to achieve the target grade of 35% manganese.

The focus of work shifted to the evaluation of reduced sulphuric acid leaching of ore and lower grade gravity concentrates. The ores and concentrates were leached successfully under conditions generally representative of conventional leaching and heap leaching. Further work was required to define a feasible process flowsheet encorporating this reducing leach step.

At a project review meeting with the Scientific Authority on October 20, 1986, a decision was reached that testwork would be terminated and that remaining monies would be directed towards an expanded scope of economic studies. This change was to ensure that any conceptual flowsheet to be investigated would have a sound economic basis. Detailed flowsheet development studies would then be persued in a Phase II Contract.

Subsequent meetings with CANMET and the New Brunswick Department of Natural Resources resulted in a specific approved scope of work, which included:

- Costs for a conventional $MnSO_4$ process.
- Costs for the production of a middling grade chemical precipitate.
- Cost effects of ore pre-concentration, electrolytic recovery, and heap leaching.

The basis of the cost evaluation was to be a production rate of

10,000 tpy contained manganese as per the original contract specification.

This scope of work was completed and results are presented in this report. A review of the developed costs showed that the analysis needed to be expanded to permit a comprehensive economic evaluation of the project. Accordingly, the effects of combining pre-concentration and heap leaching; inclusion of new technology for purification and recovery; and effects of processing with a "minimal" plant were evaluated, and are also presented in this report.

3.0 BASIS OF ESTIMATES

3.1 Production Criteria and Basis of Estimates

The capital and operating cost estimates have been prepared for alternative processes and plants to produce 10,000 tonnes per year of contained manganese in various appropriate products.

For each alternative studied, the ore tonnage has been adjusted to reflect the efficiency of the selected process. It has been assumed that the ore reserves are large enough such that processing inefficiencies have no effect on the useful life of the processing plants.

3.2 General Design Criteria

The general design criteria is based on information which is specific to the orebody and on design factors that have proven to be suitable for other projects of a generally similar nature.

Project Location	New Brunswick
Production	
Product Quantity	10,000 tonnes per year contained Mn
Product Alternatives	Medium grade $Mn(OH)_2$ Agricultural grade $MnSO_4$ High purity electrolytic Mn High purity MnO_2
Operating Schedule	330 days per year 24 hours per day

Average Tonnage	As required to meet production
Design Tonnage	Average tonnage plus 10 percent
Ore	
Analysis of ore composite	
Major MnO	9.69%, 7.57% Mn
Fe_2O_3	14.80, 9.46 Fe
SiO_2	38.70
Al_2O_3	11.0
CaO	6.65
P_2O_5	0.55
K_2O	2.59
MgO	2.50
Na_2O	0.91
TiO_2	0.49
LOI	11.30
Minor Ni	78 ppm
Co	77
As	77
Zn	76
Cu	46
Pb	4
Bi	0.9
Cd	0.2
Specific Gravity	3.2
Bulk Density	1970 kg/m^3

Analysis of a leach lixiviant

Ag	3.1 mg/L
Al	1100
As	.005
B	12.2
Ba	.06
Be	.23
Bi	11.2
C	85
Ca	537
Cd	.65
Ce	1.7
Co	7.6
Cr	5.8
Cu	7.9
Fe	5800
Hg	7.0
K	215
La	.59
Mg	475
Mn	10400
Mo	2.5
Na	13400
Nb	44
Ni	4.9
P	123
Pb	100
S	High
Sb	16.8
Se	18.5
Si	575

Sn	5.6
Sr	7.2
Te	7.5
Th	8.4
Ti	.5
U	20
V	2.2
W	5.1
Y	1.1
Zn	5.5
Zr	.9

3.3 Scope of Work and Methodology

Scope of Work

The scope of work for this economic evaluation was outined in the Project Review Report of December 11, 1986.

The scope of work included:

- Order of magnitude capital and operating costs for a standard hydrometallurgical flowsheet for the production of a "middling grade" chemical precipitate
- Effects of ore pre-concentration on capital and operating costs for the standard flowsheet
- Effects of heap leaching on capital and operating costs for the standard flowsheet
- Effects of an electrolytic recovery circuit on capital and operating costs for the standard flowsheet
- Order of magnitude capital and operating costs for a conventional hydrometallurgical flowsheet for production of agricultural grade $MnSO_4$

The above listed scope of work has been accomplished and is presented in following sections of this report. The scope of work was also expanded to include additional alternative approaches.

A brief review of the capital and operating costs of the first five process alternatives indicated that the pre-concentration flowsheet resulted in significantly lower operating costs, while the heap leaching flowsheet resulted in substantially lower capital cost. A comprehensive economic evaluation of this project would not be complete without investigation of the effects of a combination of these two process options. Accordingly this combination was evaluated.

Experience with other generally similar projects suggested that a new process as an alternative to electrowinning to produce a desirable high purity product might be developed with an intensive research and development program. Accordingly the capital and operating costs were calculated for such a postulated hydrometallurgicl process. Conversely, cost estimates were developed for a very minimal process for production of a low purity manganese sulphate to provide information on cost variations.

The five additional process alternatives selected for evaluation for the above reasons were:

- Effects of a combination of pre-concentration and heap leaching on the standard flowsheet

- Effects of a combination of pre-concentration and heap leaching on the standard flowsheet with electrowinning recovery

- Order of magnitude capital and operating costs for a hydrometallurgical flowsheet incorporating solvent extraction for recovery and purification of a chemical manganese product of high purity equivalent to an electrolytic product.

- Effects of a stripped down hydrometallurgical process on the capital and operating costs for $MnSO_4$ production

- Effects of a combination of pre-concentration and heap leaching on capital and operating costs for a conventional hydrometallurgical flowsheet for $MnSO_4$ production

These five additional alternatives are also presented in following sections of this report.

Methodology

For comparative economic studies of this nature, a common approach is to develop the economics for various alternatives from a base case. Many of the costs applicable to the base case are directly applicable to the other alternatives, thus reducing errors during the comparisons.

The above approach has been used for this study. The base case selected was the standard hydrometallurgical flowsheet for production of a middling grade chemical product. This flowsheet was selected because the process is known, is relatively complex and includes most of the unit operations that are common to the other alternatives that have been evaluated. For the alternative flowsheets, capital and operating costs were developed for those unit operations which were different or not used in the base case. These costs were then added to the costs for the base case. The costs for unit operations in the base which were not included in the alternative process flowsheet were deleted. Where necessary, the remaining unit operations in the base case were adjusted for size and costs as dictated by the requirements of the alternative process flowsheet.

3.4 Capital Cost Basis and Limits of Estimate

The capital costs developed for all the alternatives are for a grass roots, battery limits facility.

The capital costs include the following in terms of materials and labour:

- Plant site preparation
- New process equipment
- Process piping
- Process electrical
- Process instrumentation
- Mill building
- Auxiliary buildings related to processing such as analytical laboratory and warehouse
- Field services
- Engineering design, project management and purchasing
- Mobile equipment necessary for processing.

The capital costs do not include:

- Plant site purchase
- Access roads
- Feeder power lines
- Mine site preparation and development
- Mine capital equipment
- Tailings pipeline
- Tailing disposal site preparation and development
- Environmental impact statements, licences and insurance
- Corporate office expenses
- and other items not specifically listed as included.

Process equipment sizes for the capital cost calculations have been determined from and based on:

- The preliminary test program
- The material and energy balances for each alternative process evaluated
- Information in the files of Witteck Development Inc.
- Equipment sizes and sizing informaion available from recent projects of a generally similar nature
- Extensive engineering experience and prudent judgement.

The capital cost estimates are factored estimates. Total plant costs have been developed from the process equipment costs.

The process equipment costs were obtained from:

- Budget price quotations from equipment manufacturers and suppliers
- Price information in the files of Witteck Development Inc.
- Calculation on the basis of current prices for equipment fabrication
- Equipment cost for projects of a generally similar nature.

The factors used in determining the total plant cost were obtained from:

- Recognized publication on cost estimation for the Canadian mining industry
- Factor information in the files of Witteck Development Inc.
- Factors which have proven to be satisfactory for projects of a generally similar nature.

The plant capital costs have been factored from new equipment costs. The assumptions and factors used for equipment sizing are presented in detail in the Appendices.

The factors used in developing the total capital cost are as follows:

Total new equipment cost	$ A
Equipment installation 27.2% of A	0.272 A
Piping, materials & installation 27.2% of A	0.272 A
Electrical, materials & installation 27.2% of A	0.272 A
Instrumentation, materials & installation 13.0% of A	0.130 A
Main mill building, complete 54.3% of A	0.543 A
Auxilliary buildings 16.3% of A	0.163 A
Plant services, water, air, steam etc. 16.3% of A	0.163 A
Site preparation 19.6% of A	0.196 A
Field expenses 13.0% of A	0.130 A
Project engineering, management, accounting, purchasing 35.9% of A	0.359 A
Total Plant Cost	$3.50 A

The developed order of magnitude capital cost estimates are suitable for comparison purposes for evaluation of alternative processes, for planning of research programs and for deciding whether further development expenditures are reasonably justified.

The estimates are neither suitable for detailed economic evaluation of the project, nor for engineering design. The estimates should not be used for these purposes.

Within the constraints of the basis of calculation and the limitation of comparative purposes only, the accuracy of the estimates relative to each other is considered to be better than ten percent.

The estimates are of adequate accuracy to permit evaluation of the project as to potential to justify further development.

The estimates reflect general costs and not site specific costs. The costs do not include allowances for inflation.

3.5 Operating Cost Basis and Limits of Estimate

The operating cost developed for all the alternatives are for a grass roots battery limit facility.

The operating cost estimates include the following:

- Process reagents
- Direct operating labour
- Plant management and administration
- Labour overhead costs
- Electrical energy
- Fuel oil
- Operating supplies
- Maintenance labour and supplies.

The operating cost estimates do not include the following:

- Mining and related costs
- Project management and administration
- Secondary and tertiary effluent treatment
- Insurances
- Taxes and royalties
- Corporate office expenses
- and other items not specifically listed as included.

The operating costs are based on plant requirements as determined by:

- Material and energy balances
- Measured reagent requirement
- Calculated reagent requirements
- Preliminary testwork
- Process complexity
- Plant capital cost.

Unit costs were obtained from:

- Reagent manufacturers and commodity suppliers
- Price and wage information in the files of Witteck Development Inc.
- Cost information from projects of a generally similar nature.

Reagent prices at the plant site, used for this study were as follows:

Sulphuric acid	$130/tonne
Sulphur dioxide	$200/tonne
Calcium chloride	$398/tonne
Lime	$ 98/tonne
Ammonia	$410/tonne
Hydrogen sulphide	$880/tonne
Flocculant	$3100/tonne

Energy prices used were as follows:

Electrical energy	$0.05/kWh
Fuel oil	$0.258/litre

Labour costs have been developed from the following base rates:

Hour labour	$14.0/hour
Shift foremen, supervision	$37,000/year
Payroll burden	17% of salary or hourly rate

Maintenance costs were factored on the basis of plant capital cost. Maintenance costs including labour, payroll burden and maintenance supplies but excluding operating supplies are 3.5% of capital cost.

Operating supply requirements vary widely between the various alternatives studied, and are stated in detail in the Appendices. Requirements are based on generally similar projects.

The estimated operating costs are suitable for the same purposes as outlined for the capital costs. The accuracy of estimates relative to each other is better than ten percent. The estimates are of adequate accuracy to permit a project justification review.

4.0 CAPITAL AND OPERATING COSTS FOR ALTERNATIVE PROCESSES

4.1 Standard Hydrometallurgical Process for Production of a Middling Chemical Product, $Mn(OH)_2$

Process and Plant Summary

The hydrometallurgical process selected to represent the base study case is outlined in Figure No. 1. To Witteck's knowledge, this process has never been used commercially, however it encompasses standard hydrometallurgical process technology. Similar conceptual flowsheets have been published (1) as early as 1950. See "Manganese" by A. H. Sully, page 94 in Appendix XI.

In a process plant, run of mine ore would be reduced in a primary crusher and ground in a conventional SAG-ball mill circuit to 80% -200 mesh. The ground ore is then thickened, filtered and repulped with a portion of the pregnant solution to 50% solids.

Leach conditions are pH 2.0 and 80°C. Leaching takes place for 24 hours in agitated leach tank-gas reactors with sulphuric acid and sulphur dioxide gas addition. The sulphur dioxide maintains reducing conditions. Abour 80% of the manganese is extracted to solution under these conditions.

Leach discharge is filtered and washed in two stages for soluble manganese recovery. The residue at 20% moisture is advanced to tailings treatment for neutralization with lime.

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(1) Hoak, R. D. and Coull, J. Chem. Engng. Progr. 46/3 (1950) 158.

The recovered pregnant solution is treated with recycled calcium chloride solution for removal of sulphate as calcium sulphate:

$$MnSO_4 + CaCl_2 \longrightarrow CaSO_4\downarrow + MnCl_2$$

Lime slurry is then added to raise the pH from 2.0 to 4.5 to precipitate impurities such as iron and aluminum. The solids are removed by thickening and filtration and discharged to the tailings treatment circuit.

The purified pregnant solution is advanced to the manganese recovery circuit. Here, lime slurry is added to pH 9.5 to precipitate manganese hydroxide, i.e:

$$MnCl_2 + Ca(CH)_2 \longrightarrow Mn(OH)_2\downarrow + CaCl_2$$

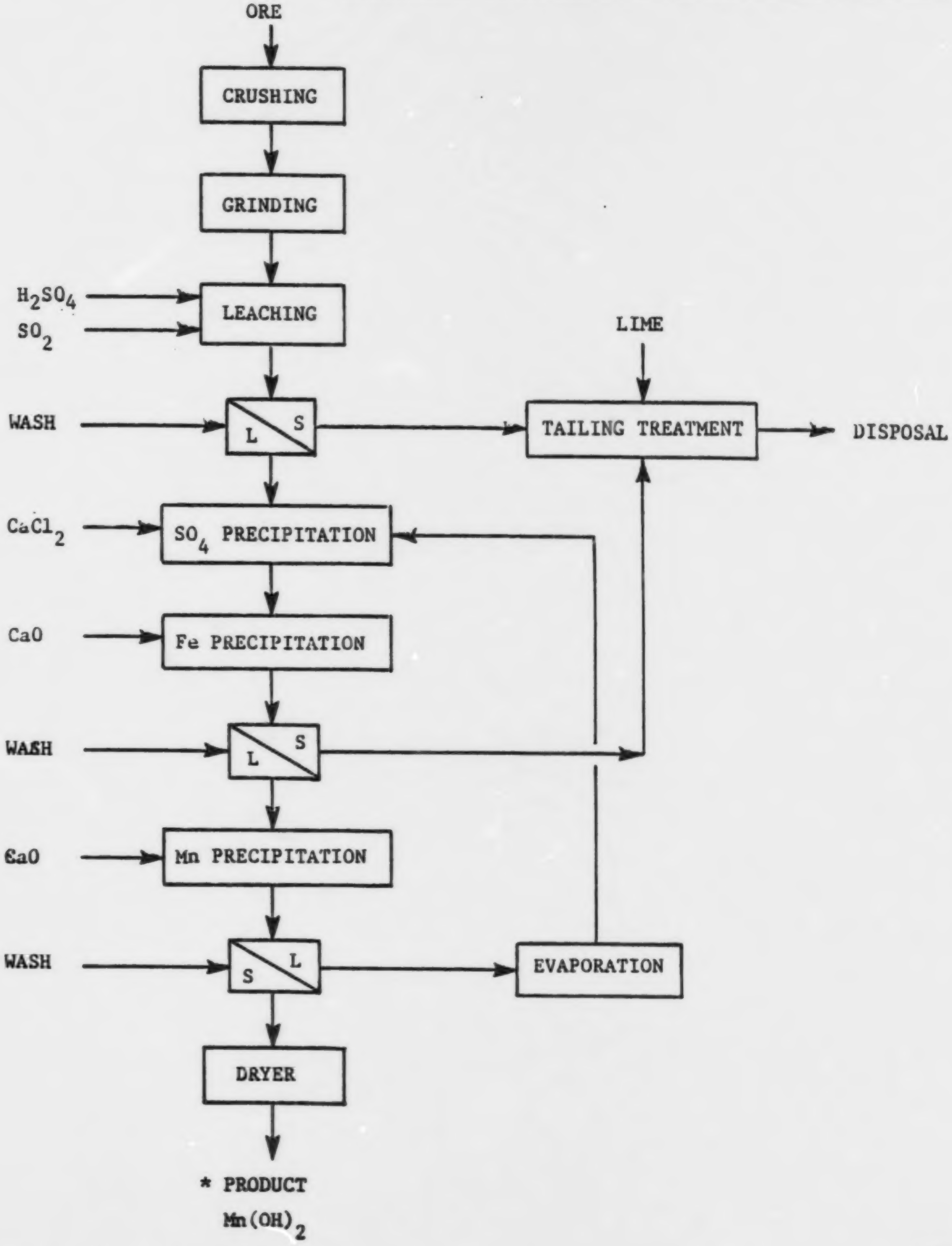
Use of the $MnCl_2$ intermediary results in greatly improved product purity due to the absence of gypsum contamination.

The precipitated product is thickened, filtered to 50% moisture and washed prior to drying in a multi-hearth dryer. The dried product is stored in a bin for shipment via tanker truck. The depleted chloride solution is concentrated in a multi-effect evaporation system for recovery and concentration of calcium chloride solution, which is recycled. A $CaCl_2$ loss of 5% is expected. Overall manganese recovery in the plant is 76.8%.

A more detailed process description is provided in Appendix I.

Figure 1

HYDROMETALLURGICAL BASE CASE PROCESS FLOWSHEET
PRODUCTION OF A "MIDDLING GRADE" CHEMICAL PRECIPITATE



ORE
CRUSHING
GRINDING
H_2SO_4
SO_2
LEACHING
LIME
WASH
L
S
TAILING TREATMENT
DISPOSAL
$CaCl_2$
SO_4 PRECIPITATION
CaO
Fe PRECIPITATION
WASH
L
S
CaO
Mn PRECIPITATION
WASH
S
L
EVAPORATION
DRYER
* PRODUCT
$Mn(OH)_2$

* Can be calcined to MnO_2 ± 60% Mn (Max Mn Content 63.2%)

Design Criteria

The basic design criteria for this alternative process, the base case, are presented below. Additional design criteria for equipment sizing and plant design are shown in Appendix I.

Ore Tonnage	- annual	172,000 tonne
	- average	21.70 tonne/h
	- design	23.90 tonne/h
Ore Grade		7.57% Mn
Production	- annual	10,000 tonne contained Mn
Product		$Mn(OH)_2$, 60% Mn
Recovery	- leaching	80.0%
	- overall	76.8%
Pregnant Solution Flow to Purification		36.95 tonne/h
Acid Consumption		180 kg/tonne ore
SO_2 Consumption		90 kg/tonne ore
Lime Consumption		115 kg/tonne ore
$CaCl_2$ Consumption		10.7 kg/tonne ore

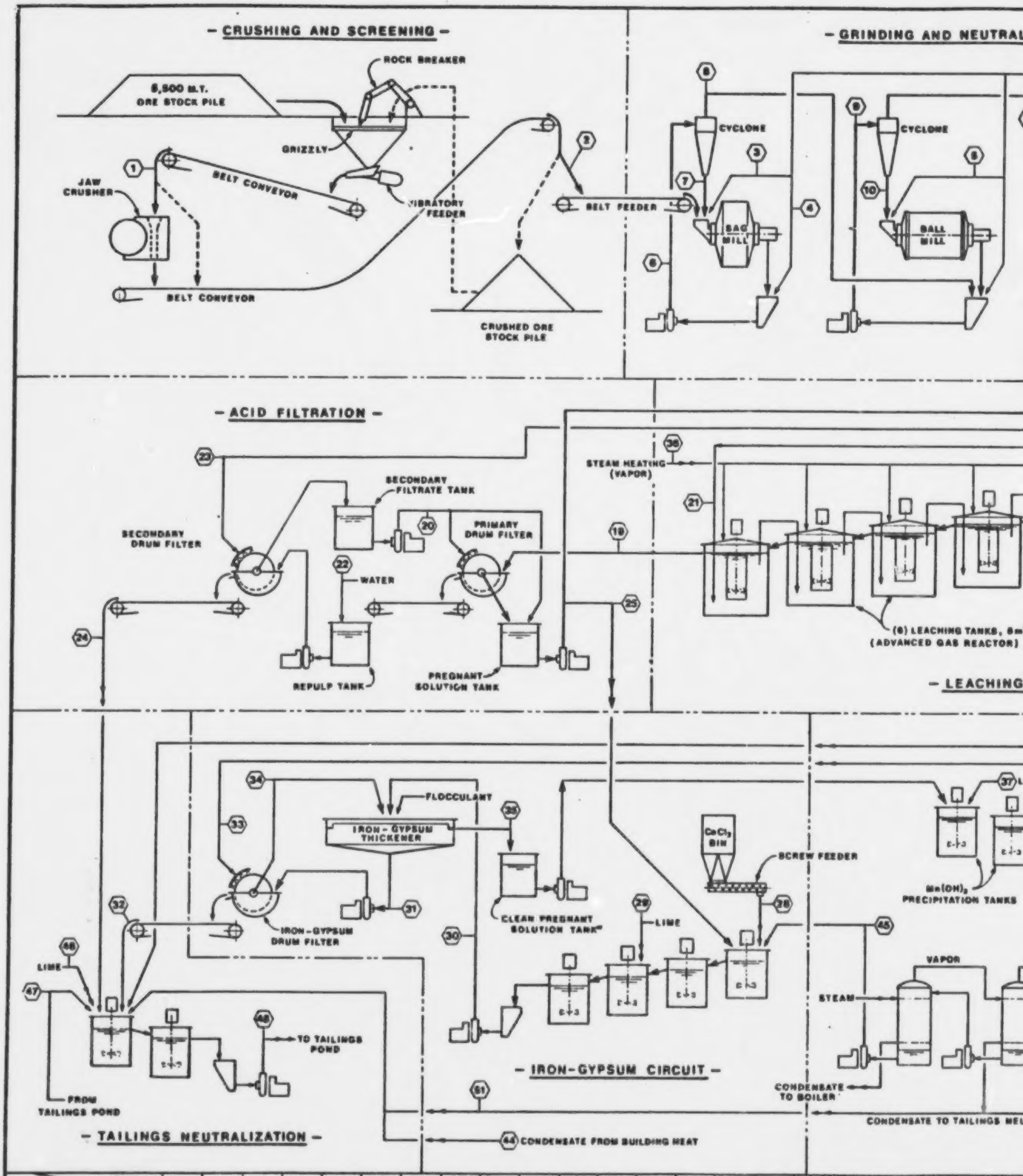
Flowsheet - Material and Energy Balance

The flowsheet for this process alternative is presented in pictorial form with equipment type in Drawing Number 100-110-001, Revision P1. The material balance is shown on the drawing.

The volumetric flowrates are approximate and are suitable for sizing process equipment.

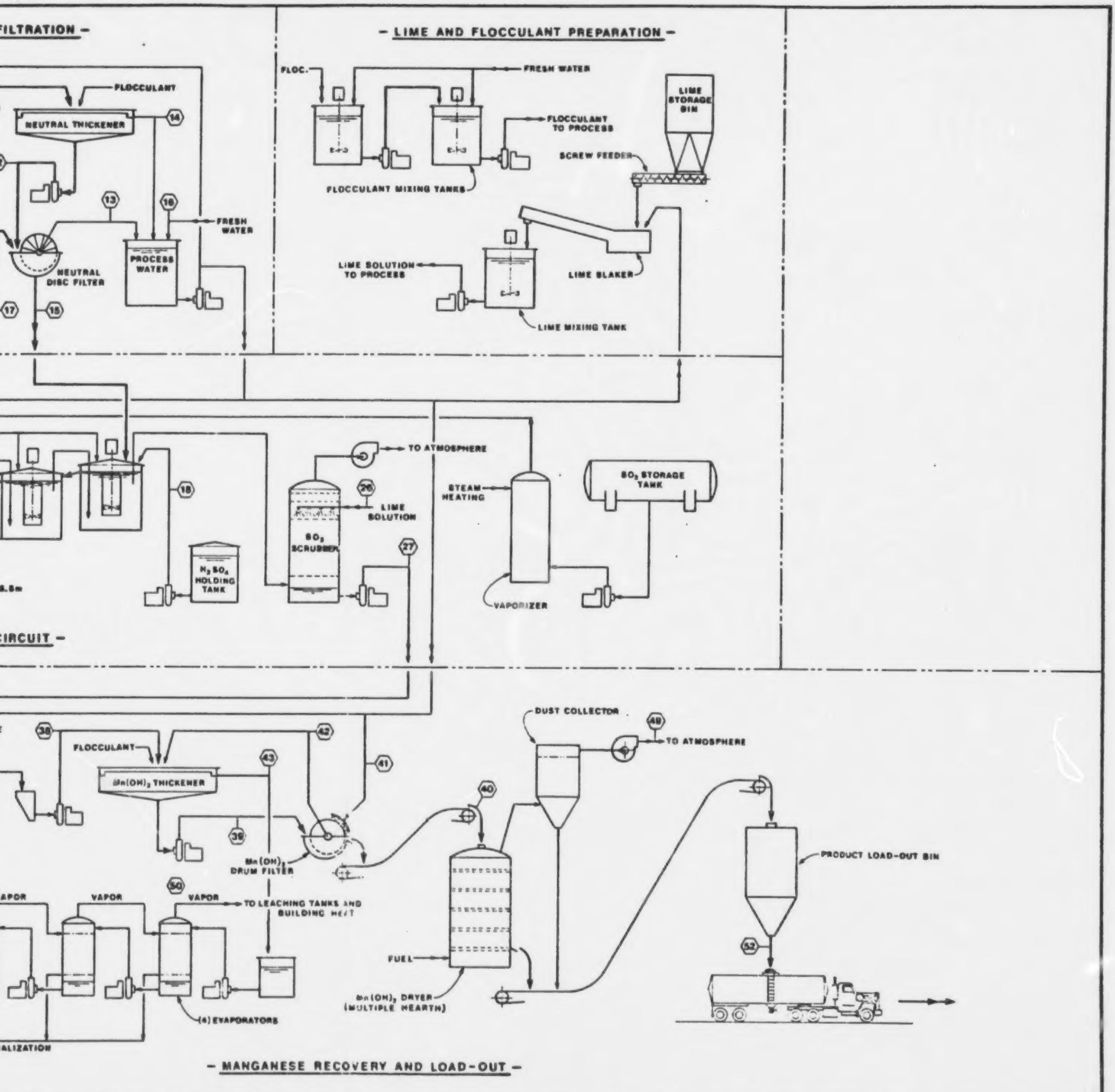
Energy requirements were calculated to be as follows:

Electrical energy	10.2×10^6 kWh/year
Heat	39.6×10^6 BTU/h


- CRUSHING AND SCREENING -
6,500 M.T. ORE STOCK PILE
ROCK BREAKER
GRIZZLY
BELT CONVEYOR
VIBRATORY FEEDER
JAW CRUSHER
BELT CONVEYOR
BELT FEEDER
CRUSHED ORE STOCK PILE
1
2
- GRINDING AND NEUTRAL
CYCLONE
SAG MILL
CYCLONE
BALL MILL
3
4
5
6
7
8
9
10
- ACID FILTRATION -
SECONDARY FILTRATE TANK
SECONDARY DRUM FILTER
PRIMARY DRUM FILTER
WATER
REPULP TANK
PREGNANT SOLUTION TANK
19
20
22
23
24
25
STEAM HEATING (VAPOR)
36
21
(6) LEACHING TANKS, 8m
(ADVANCED GAS REACTOR)
- LEACHING
FLOCCULANT
IRON - GYPSUM THICKENER
IRON-GYPSUM DRUM FILTER
CLEAN PREGNANT SOLUTION TANK
CaCl₂ BIN
SCREW FEEDER
LIME
26
29
30
31
32
33
34
35
37
Mn(OH)₂ PRECIPITATION TANKS
45
VAPOR
STEAM
CONDENSATE TO BOILER
CONDENSATE TO TAILINGS NEU
LIME
46
47
48
TO TAILINGS POND
FROM TAILINGS POND
- IRON-GYPSUM CIRCUIT -
51
44 CONDENSATE FROM BUILDING HEAT
- TAILINGS NEUTRALIZATION -

DESCRIPTION \ STREAM		1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22
FLOW	SOLIDS TPH	35.85	23.90	—	—	—	59.75	35.85	23.90	59.75	35.85	23.90	23.90	—	—	23.90	—	—	—	20.32	—	—	—
	SOLUTION TPH	3.98	2.66	1.34	99.87	0.82	139.42	35.85	103.87	139.42	29.33	110.00	23.90	17.93	86.19	14.30	49.27	8.48	4.30	27.44	23.04	2.18	17.9
	VOLUME m³/Hr	—	—	1.3	100	0.8	150	47	111	158	40	118	31	18	86	22	49	8.5	3.4	26	23	—	18
SLURRY DENSITY %		90	90	—	—	—	30	50	18.9	30	55	17.8	50	—	—	62.6	—	—	—	42.5	—	—	—

DWG. No.	REFERENCE	DWG. No.	REFERENCE	No.	DATE	REVISIONS


FILTRATION -
FLOCCULANT
NEUTRAL THICKENER
14
13
16
FRESH WATER
NEUTRAL DISC FILTER
PROCESS WATER
17
15
- LIME AND FLOCCULANT PREPARATION -
FLOC.
FRESH WATER
LIME STORAGE BIN
FLOCCULANT TO PROCESS
SCREW FEEDER
FLOCCULANT MIXING TANKS
LIME SOLUTION TO PROCESS
LIME SLAKER
LIME MIXING TANK
TO ATMOSPHERE
18
26
LIME SOLUTION
SO_2 SCRUBBER
27
H_2SO_4 HOLDING TANK
STEAM HEATING
SO_2 STORAGE TANK
VAPORIZER
CIRCUIT -
38
FLOCCULANT
42
43
41
$Mn(OH)_2$ THICKENER
39
$Mn(OH)_2$ DRUM FILTER
40
DUST COLLECTOR
49
TO ATMOSPHERE
PRODUCT LOAD-OUT BIN
VAPOR
VAPOR
50
VAPOR
TO LEACHING TANKS AND BUILDING HEAT
FUEL
$Mn(OH)_2$ DRYER (MULTIPLE HEARTH)
52
(4) EVAPORATORS
ALIZATION
- MANGANESE RECOVERY AND LOAD-OUT -

23	24	25	26	27	28	29	30	31	32	33	34	35	36	37	38	39	40	41	42	43	44	45	46	47	48	49	50	51	52
—	30.32	—	0.24	0.52	0.265	1.52	9.73	9.73	9.73	—	—	—	—	1.07	2.25	2.25	2.25	—	—	—	—	—	0.01	—	30.58	—	—	—	2.25
5.06	5.06	36.98	0.75	0.69	—	4.56	46.56	38.92	9.73	9.73	38.92	46.56	3.23	5.61	50.76	9.0	2.25	2.25	9.0	50.76	0.66	12.0	0.03	5.70	56.76	2.25	3.88	34.9	—
5	11	37	0.8	0.9	—	5	50	43	14	10	39	46	—	6.4	51	9.5	2.5	2.3	9.0	50	6.7	12.0	0.1	5.7	66	—	—	35	—
—	60	—	25	46	—	25	17.5	20	50	—	—	—	—	25	4.2	20	50	—	—	—	—	—	25	—	35	—	—	—	—

	BY	CHECK	No.	DATE	REVISIONS	BY	CHECK

SCALE	N/A	
DRAWN	C.F. Krause	06 Feb.'87
CHECKED		
APPROVAL		

WDI

WITTECK DEVELOPMENT INC.

DEPARTMENT OF SUPPLY AND SERVICES CANADA
NEW BRUNSWICK MANGANESE MINERAL DEPOSITS
EVALUATION OF PROCESS OPTIONS
HYDROMETALLURGICAL FLOWSHEET - BASE CASE

DATE | PROJECT No 52°1 | DWG No 100-110-001 | REV P 1

Capital Cost Estimate

The capital cost for this process alternative, the base case, has been estimated as follows:

Crushing and grinding	$5,659,000
Neutral thickening and filtration	1,362,000
Leaching	3,042,000
Acid filtration	3,759,000
Iron-gypsum precipitation	3,199,000
Manganese precipitation; reagent recovery	7,318,000
Tailing neutralization	452,000
Reagent storage and distribution	1,358,000
Process air, vacuum and steam	1,624,000
Total Plant Cost	27,773,000
Mobile equipment	544,000
Contingency, 20%	5,554,000
Total Plant Capital Cost Estimate	$33,871,000

Operating Cost Estimate

The operating cost for this process alternative has been estimated as follows:

	Per Year
Reagents	$ 9,924,000
Labour	2,228,000
Electrical energy	508,000
Fuel oil	2,426,000
Operating supplies	200,000
Maintenance	1,185,000
Total Operating Cost	16,471,000
Contingency, 10%	1,647,000
Total Plant Operating Cost Estimate	$18,118,000
Cost per kg Mn contained in product	$1.81

4.2 Effect of Pre-concentration on the Standard Hydrometallurgical Process

Process and Plant Summary

Run of mine ore is crushed and then ground in a SAG mill circuit to a nominal 20 mesh. This grind is suitable for gravity concentration using spiral concentrators such as Reichert cones.

About 50% of the manganese is recovered and concentrated in 30% of the original ore weight. This may appear inefficient, however sufficient acid consuming gangue minerals are rejected to lower leach reagent requirements by 50%. The concentrate manganese grade is double the ore grade.

The spiral rejects are discharged to tailings. The concentrate is ground further and treated in the same circuits as utilized for the standard (base case) flowsheet. The sizes of the various circuits have been adjusted as necessary to compensate for the changes in flowrates resulting from the change in feed tonnage to the circuit.

The sizes of the product drying and handling circuits remain unchanged since the quantity of product produced remains unchanged.

A detailed description of the pre-concentration process and its effects on the overall flowsheet is presented in Appendix II.

Design Criteria

The basic design criteria for the pre-concentration option for the standard hydrometallurgical flowsheet are presented below. Additional design criteria for engineering calculations are shown in Appendix II.

Ore Tonnage	- annual	325,207 tonne
	- average	41.01 tonne/h
	- design	45.16 tonne/h
Pre-concentration		
Design	- feed	45.16 tonne/h
	- concentrate	12.41 tonne/h
	- tailing	32.75 tonne/h
Recovery	- manganese	52.28%
	- weight	27.48%
Production	- annual	10,000 tonne contained Mn
Product		$Mn(OH)_2$, 60% Mn
Recovery from Concentrate -		
	- leaching	80.0%
	- overall	76.8%
Pregnant Solution Flow to Purification		18.6 tonne/h
Acid Consumption	- concentrate	180 kg/tonne
	- ore basis	49.4 kg/tonne ore
SO_2 Consumption	- concentrate	90 kg/tonne
	- ore basis	24.7 kg/tonne ore
Lime Consumption	- concentrate	181 kg/tonne
	- ore basis	49.6 kg/tonne ore
$CaCl_2$ Consumption	- concentrate	18.9 kg/tonne
	- ore basis	5.2 kg/tonne ore

Flowsheet; Material and Energy Balance

The flowsheet for the pre-concentration circuit is shown in Figure No. 2. The relevant material balance details are presented in Appendix II.

Energy requirements for the complete standard hydrometallurgical flowsheet incorporating pre-concentration have been calculated to be as follows:

Electrical energy	10.4×10^6 kWh/year
Heat	23.8×10^6 BTU/h

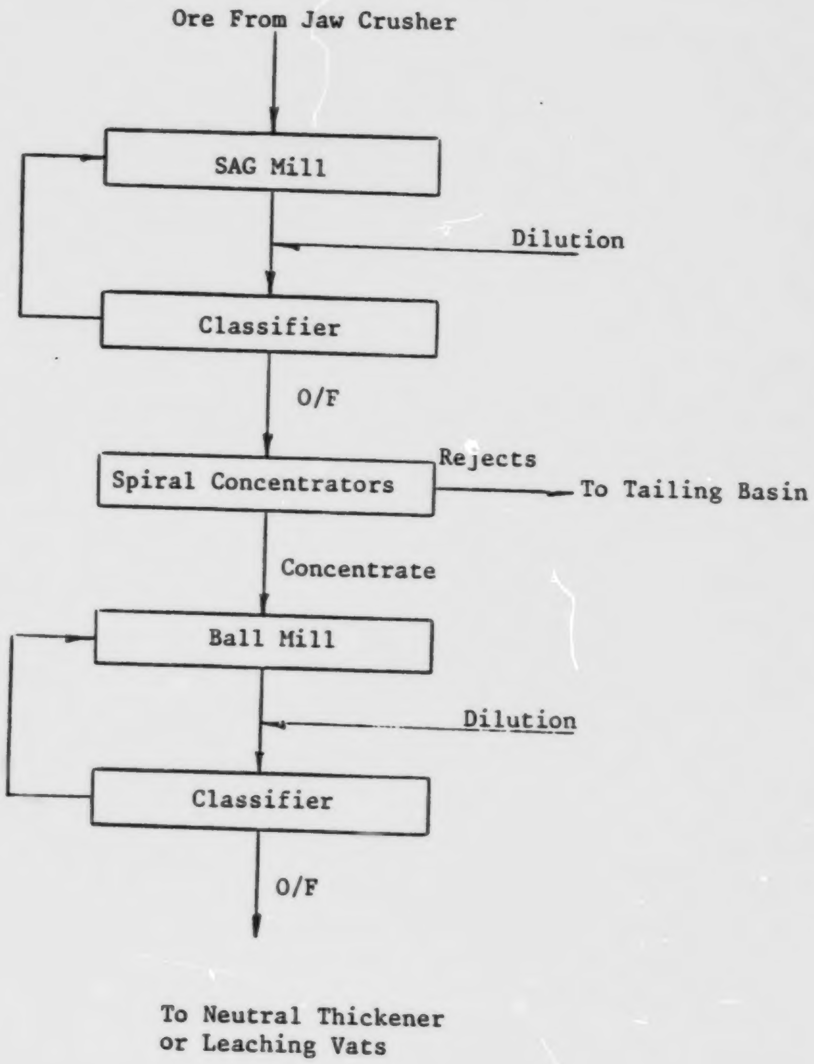
Capital Cost Estimate

The capital cost for the standard hydrometallurgical flowsheet with pre-concentration has been estimated as follows:

Crushing and grinding	$ 6,510,000
Spiral concentration	473,000
Neutral thickening and filtration	919,000
Leaching	2,053,000
Acid filtration	2,537,000
Iron-gypsum precipitation	2,289,000
Manganese precipitation; reagent recovery	6,041,000
Tailing neutralization	341,000
Reagent storage and distribution	1,358,000
Process air, vacuum and steam	1,386,000
Total Plant Cost	23,907,000
Mobile equipment	544,000
Contingency, 20%	4,781,000
Total Plant Capital Cost Estimate	$29,232,000

Figure No. 2

PRE-CONCENTRATION PROCESS FLOWSHEET


Ore From Jaw Crusher
SAG Mill
Dilution
Classifier
O/F
Spiral Concentrators
Rejects
To Tailing Basin
Concentrate
Ball Mill
Dilution
Classifier
O/F
To Neutral Thickener
or Leaching Vats

Operating Cost Estimate

The operating cost for the standard hydrometallurgical flowsheet incorporating pre-concentration has been estimated to be:

	Per Year
Reagents	$ 6,031,000
Labour	2,389,000
Electrical energy	517,000
Fuel oil	1,470,000
Operating supplies	339,000
Maintenance	1,023,000
Total Operating Cost	11,769,000
Contingency, 10%	1,177,000
Total Plant Operating Cost Estimate	$12,946,000
Cost per kg Mn contained in product	$1.30

4.3 Effect of Heap Leaching on the Standard Hydrometallurgical Process

Process and Plant Description

Manganese extraction by heap leaching results in the deletion of the grinding, acid leaching and acid filtration circuits of the base case.

Run of mine ore is crushed to 3/8 inch in a three stage crushing and screening plant. The sized ore is trucked to the prepared leach pads and stacked by a large front-end loader. The pads are leached in tandem with solution from the oldest pad, in its last 20 days of the total 40 day leach period advanced to the newest pad in its initial 20 day leach period. This arrangement produces a relatively constant grade of leach solution for purification and manganese recovery. The volume of leach solution bleed to purification and recovery is equal to the volume of pregnant solution produced in the standard hydrometallurgical process.

The three stage crushing circuit and the heap leaching process replace the crushing and grinding circuit, the neutral thickening and filtration circuit, the agitator leaching circuit, and the two stage acid filtration circuit used in the standard hydrometallurgical process.

The heaps are washed and left as permanent tailing disposal heaps after the completion of the leaching cycle. New leaching pads are constructed and commissioned as required by the production schedule. Details of this process are presented in Appendix III.

Design Criteria

The basic design criteria for the heap leaching option for the standard hydrometallurgical flowsheet are presented below. The design criteria necessary for engineering calculations for the heap leach option are presented in Appendix III.

Ore Tonnage	-annual	209,683 tonne
	-average	26.48 tonne/h
	-design	29.12 tonne/h
Ore Grade		7.57% Mn
Production	- annual	10,000 tonne contained Mn
Product		$Mn(OH)_2$, 60% Mn
Leaching Period		-primary 20 days
	-secondary	20 days
	-total	40 days
Recovery	-leaching	65.0%
	-overall	63.0%
Leach Solution Bleed to Process		36.95 tonne/h
Acid Consumption		180 kg/tonne ore
SO_2 Consumption		90 kg/tonne ore
Lime Consumption		107 kg/tonne ore
$CaCl_2$ Consumption		10.6 kg/tonne ore

Flowsheet; Material & Energy Balance

The flowsheet for the fine crushing and heap leaching circuits is shown in Figure No. 3. The material balance details are presented in Appendix III.

Energy requirements for the standard hydrometallurgical flowsheet incorporating fine crushing and heap leaching have been calculated as follows:

Electrical energy	6.4×10^6 kWh/year
Heat	41.5×10^6 BTU/h

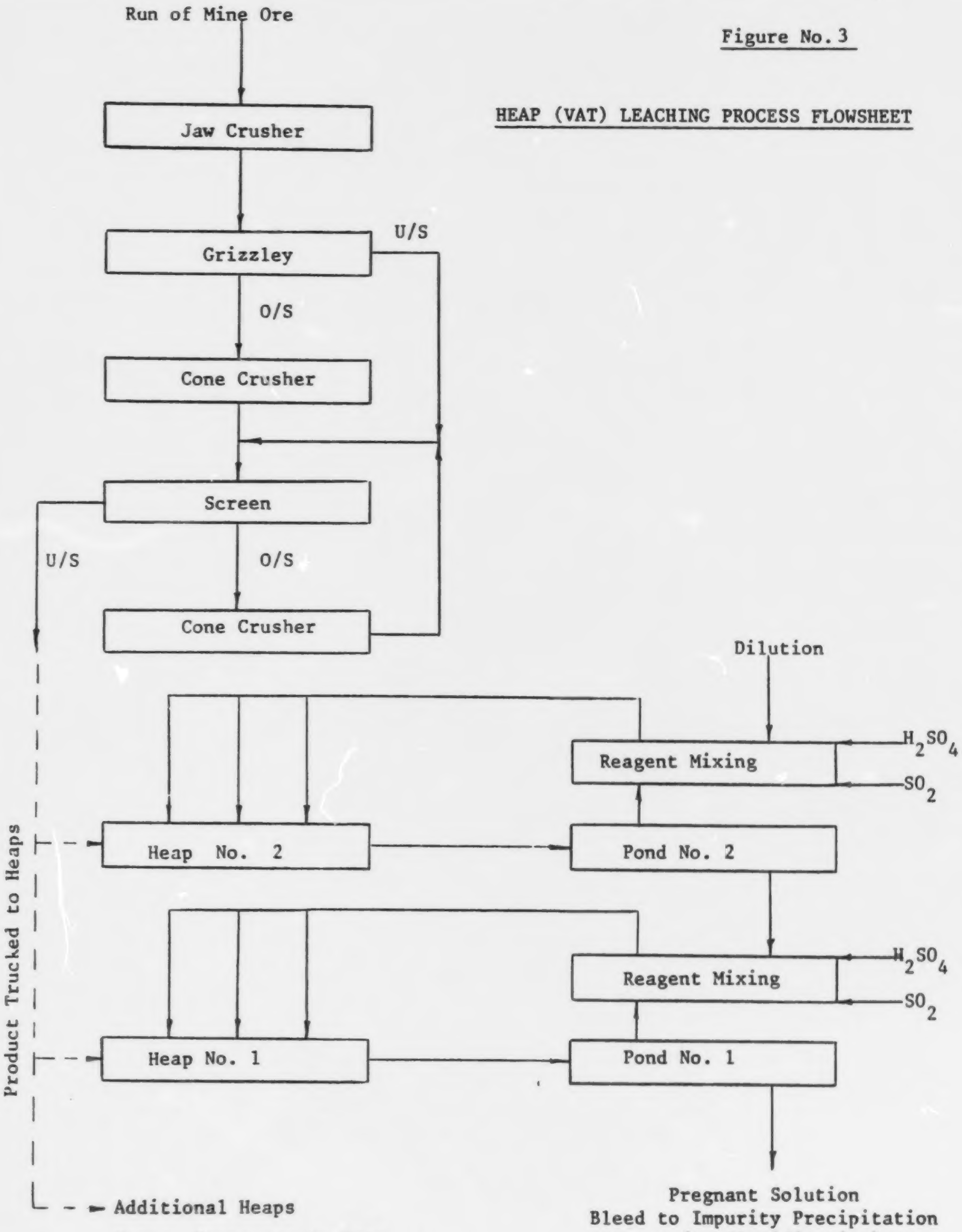
Capital Cost Estimate

The capital cost for the standard hydrometallurgical flowsheet with heap leaching has been estimated as follows:

Crushing	$2,720,000
Initial leach pads and ponds	456,000
Solution handling	527,000
Iron-gypsum precipitation	3,381,000
Manganese precipitation; reagent recovery	7,403,000
Tailings neutralization	263,000
Reagent storage & distribution	1,358,000
Process air, vacuum & steam	1,005,000
Total Plant Cost	$17,113,000
Mobile equipment	1,580,000
Contingency, 20%	3,423,000
Total Plant Capital Cost Estimate	$22,116,000

Figure No. 3

HEAP (VAT) LEACHING PROCESS FLOWSHEET


Run of Mine Ore
Jaw Crusher
Grizzley
U/S
O/S
Cone Crusher
Screen
U/S
O/S
Cone Crusher
Dilution
Reagent Mixing
H_2SO_4
SO_2
Heap No. 2
Pond No. 2
Reagent Mixing
H_2SO_4
SO_2
Heap No. 1
Pond No. 1
Product Trucked to Heaps
Additional Heaps
Pregnant Solution
Bleed to Impurity Precipitation
(or Clarification)

Note: Heaps can be Vats

Operating Cost Estimate

The operating cost for the standard hydrometallurgical flowsheet using heap leaching has been estimated to be:

Per Year	
Reagent	$10,998,000
Labour	1,992,000
Electrical energy	318,000
Fuel oil	2,601,000
Operating supplies	49,000
Leach pad preparation	708,000
Maintenance	774,000
Total Operating Cost	$17,440,000
Contingency, 10%	1,744,000
Total Plant Operating Cost Estimate	$19,184,000
Cost per kg Mn contained in product	$1.92

4.4 Effect of Electrowinning on the Standard Hydrometallurgial Process.

Process and Plant Description

Manganese is the most difficult of the common elements to electrowin from aqueous solutions due to its extreme position in the electromotive series. Meticulous purification procedures are required to achieve even low current efficiency levels. Canvass diaphragms between the anode and cathode cell compartments must be used to limit hydrogen evolution.

In spite of the technical risk and operating difficulties associated with electrowinning manganese, this option has been included to investigate the project cost advantage of a high purity product (99.98% Mn). In the resulting flowsheet, the process circuits are the same as those for the standard hydrometallurgical flowsheet from crushing and grinding through to acid filtration. However, recycle anolyte is used for repulping leach feed. The recovered pregnant solution undergoes purification for iron and base metal removal. Hydrogen sulfide is used for base metal precipitation.

The purified solution is passed through electrolytic cells where manganese metal is deposited on stainless steel cathodes. The spent electrolytic solution (anolyte) flows through cooling towers and is then recycled to the cells for temperature control and to the leaching circuit for re-use of reagents. The deposited metal is stripped off the cathodes and is crushed for sale as final product.

A detailed description of the electrowinning circuit is presented in Appendix IV.

Design Criteria

The basic design criteria for the standard hydrometallurgical process with electrolytic recovery of manganese metal are shown below. Additional design criteria for equipment sizing and engineering calculations are shown in Appendix IV.

Ore Tonnage	- annual	172,000 tonne
	- average	21.70 tonne/h
	- design	23.90 tonne/h
Ore Grade		7.57% Mn
Production	- annual	10,000 tonne Mn as manganese metal
Product		Manganese metal, 99% Mn
Recovery	- leaching	80.0%
	- overall	76.8%
Pregnant Solution Flow to Purification		36.96 tonne/h
Electrowinning Power Requirement		9.9 kWh/kg Mn
Acid Consumption		180 kg/tonne ore
SO_2 Consumption		90 kg/tonne ore
Lime Consumption		143 kg/tonne ore

Flowsheet; Material & Energy Balance

The flowsheet for the electolytic circuit for production of manganese metal is presented in Figure No 4. The material balance details are presented in Apprendix IV.

Energy requirements for the complete electrolytic flowsheet were calculated to be as follows:

Electrical energy	115.2×10^6 kWh/year
Heat	19.1×10^6 BTU/h

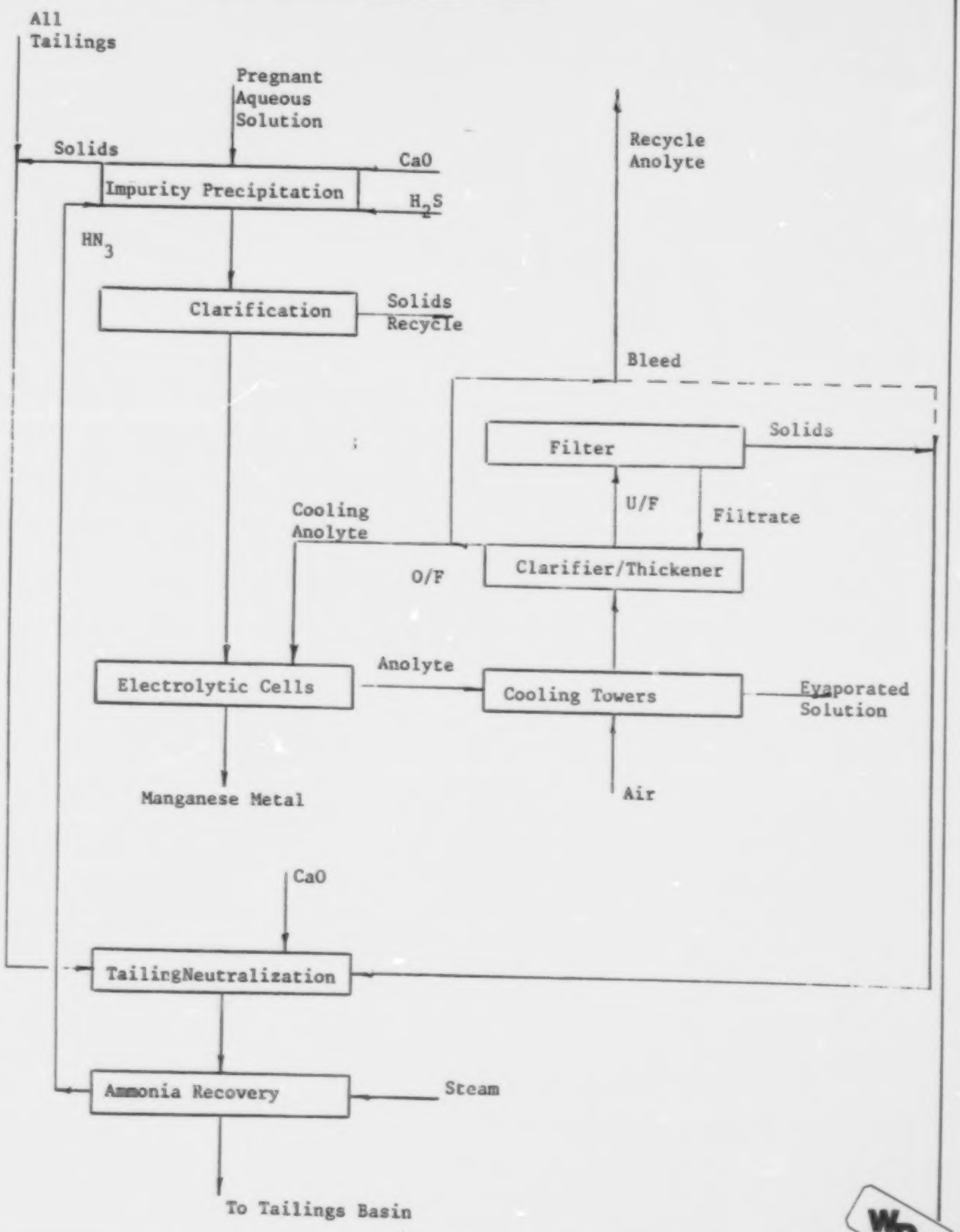
Capital Cost Estimate

The capital cost for the standard hydrometallurgical flowsheet with modifications to include electolytic recovery of manganese metal has been estimated as follows:

Crushing & grinding	$5,659,000
Neutral thickening & filtration	1,362,000
Leaching	3,042,000
Acid filtration	3,759,000
Impurity purification	3,199,000
Clarification	809,000
Electrowinning	13,731,000
Anolytic treatment	7,767,000
Tailing neutralization	879,000
Reagent storage & distribution	1,985,000
Process air, vacuum & steam	1,386,000
Total Plant Cost	43,578,000
Mobile Equipment	577,000
Contingency, 20%	8,716,000
Total Plant Capital Cost Estimate	$52,871,000

Figure No. 4

ELECTROWINNING PROCESS FLOWSHEET


All Tailings
Pregnant Aqueous Solution
Solids
CaO
Impurity Precipitation
H_2S
HN_3
Recycle Anolyte
Clarification
Solids Recycle
Bleed
Filter
Solids
U/F
Filtrate
Cooling Anolyte
O/F
Clarifier/Thickener
Electrolytic Cells
Anolyte
Cooling Towers
Evaporated Solution
Manganese Metal
Air
CaO
TailingNeutralization
Ammonia Recovery
Steam
To Tailings Basin

Operating Cost Estimate

The operating cost for this process alternative has been estimated as follows:

	Per Year
Reagents	$ 8,672,000
Labour	3,526,000
Electrical energy	5,760,000
Fuel oil	1,186,000
Operating supplies	662,000
Maintenance	1,851,000
Total Operating Cost	21,657,000
Contingency, 10% of Total Oper.Cost	2,166,000
Total Plant Operating Cost Estimate	$23,823,000
Cost per kg of manganese metal	$2.38

4.5 Effect of Pre-concentration and Vat Leaching on the Standard Hydrometallurgical Flowsheet

Process and Plant Description

A brief review of the capital and operating costs of the various options indicated that the pre-concentration flowsheet resulted in significantly lower operating costs, while the heap leaching flowsheet resulted in substantially lower capital cost. This warranted the inclusion of these two alternatives into the standard hydrometallurgical flowsheet.

Run-of-mine ore is ground to 20 mesh in a SAG mill circuit and passed through spiral concentrators. The concentrate is pumped directly to permanent vats for percolation leaching. Rejects are discharged to tailings. The leach solution bleed is processed through the purification and precipitation circuits in the standard hydrometallurgical process. Circuit sizes have been adjusted as necessary to compensate for flow changes.

Process details are presented in Appendix IV.

Design Criteria

The basic design criteria for the standard hydrometallurgical process with pre-concentration and vat leaching is shown below.

Ore Tonnage	- annual	401,077 tonne/year
	- average	50.64 tonne/h
	- design	55.71 tonne/h
Ore Grade		7.57% Mn

Pre-concentration

Design	- feed	55.71 tonne/h
	- concentrate	15.31 tonne/h
	- tailing	40.40 tonne/h
	- recovery,manganese	52.28%
	weight	27.48%
Production	- annual	10,000 tonne contained Mn
Product		$Mn(OH)_2$, 60% Mn
Recovery from Concentrate		
	- leaching	65.0%
	- overall	63.0%
Leach Solution Bleed to Process		18.6 tonne/h
Acid consumption	- concentrate	180 kg/tonne
	- ore basis	49.4 kg/tonne ore
SO_2 consumption	- concentrate	90 kg/tonne
	- ore basis	24.7 kg/tonne ore
Lime consumption	- concentrate	174 kg/tonne
	- ore basis	47.8 kg/tonne ore
$CaCl_2$ consumption	- concentrate	17.2 kg/tonne
	- ore basis	4.7 kg/tonne

Flowsheet: Material and Energy Balance

The flowsheet for this circuit showing the combination of pre-concentration and vat leaching is a combination of the flowsheets as shown on Figure No. 2, the pre-concentration flowsheet and Figure No. 3, the heap (vat) leaching flowsheet. Material balance details are presented in Appendix V.

Energy requirements for the complete standard hydrometallurgical flowsheet incorporating pre-concentration and vat leaching have been calculated to be as follows:

Electrical energy	9.3×10^6 kWh/year
Heat	23.8×10^6 BTU/h

Capital Cost Estimate

The capital cost for the standard hydrometallurgical flowsheet incorporating pre-concentration and vat leaching has been estimated as follows:

Crushing and grinding	$ 6,188,000
Spiral concentration	536,000
Initial leach vats and ponds	486,000
Solution handling	527,000
Iron-gypsum precipitation	2,433,000
Manganese precipitation; reagent recovery	6,041,000
Tailing neutralization	234,000
Reagent storage and distribution	1,358,000
Process air, vacuum and steam	767,000
Total Plant Cost	18,570,000
Mobile equipment	544,000
Contingency, 20%	3,714,000
Total Plant Capital Cost Estimate	$22,828,000

Operating Cost Estimate

The operating cost for this process alternative has been estimated as follows:

	Per Year
Reagents	7,232,000
Labour	1,992,000
Electrical energy	470,000
Fuel oil	1,470,000
Operating supplies	305,000
Leach pad preparation	423,000
Maintenance	799,000
Total Operating Cost	12,691,000
Contingency, 10%	1,269,000
Total Plant Operatiang Cost Estimate	$13,960,000
Cost per kg Mn contained in product	$1.40

4.6 Pre-concentration and Vat Leaching for the Standard Hydrometallurical Flowsheet with Electrowinning

Process and Plant Description

The capital and operating cost advantages of vat leaching and pre-concentration have been included into the standard hydrometallurgical flowsheet with electrowinning recovery of manganese metal.

Run of mine ore is ground to 20 mesh in a SAG mill and passed through spiral concentrators. The concentrate is pumped directly into prepared vats for percolation leaching. Rejects are dischaged to tailings. The leach solution bleed is processed through purification, clarification and passed through the tank house for electrolytic recovery of manganese metal. Anolyte is processed in the anolyte treatment circuit where a portion is evaporated during solution cooling. The low moisture content of the coarsely ground and well drained concentrate permits total recycle of anolyte to leach with significant savings in acid. Process details are presented in Appendix VI.

Design Criteria

The basic design criteria for the standard hydrometallurgical process which includes pre-concentration, vat leaching and electrowinning is shown below:

Ore Tonnage	- annual	401,077 tonne/year
	- average	50.64 tonne/h
	- design	55.71 tonne/h
Pre-concentration Design	- feed	55.71 tonne/h
	- concentrate	15.31 tonne/h
	- tailings	40.40 tonne/h
	- recovery, manganese	52.28%
	weight	27.48%

Production	- annual	10,000 tonne contained Mn
Product		Mn metal, 99% Mn
Recovery from Concentrate	- leaching	65.0%
	- overall	63.0%
Leach Solution Bleed to Process		36.95 tonne/h
Electrowinning Power Requirement		9.9 kWh/kg Mn
Acid Consumption	- concentrate	180 kg/tonne
	- ore basis	49.4 kg/tonne ore
SO_2 Consumption	- concentrate	90 kg/tonne
	- ore basis	24.7 kg/tonne ore
Lime Consumption	- concentrate	174 kg/tonne
	- ore basis	47.82 kg/tonne

Flowsheet; Material & Energy Balance

The flowsheet for the pre-concentration, vat leaching and electrolytic recovery flowsheet is a combination of the flowsheets shown in Figures No. 2, No. 3 and No. 4. These processes replace the grinding, leaching, purification, product precipitation and drying circuits for the standard flowsheet.

The material balance details for this process alternative are presented in Appendix VI.

Energy requirements for the complete standard hydrometallurgical flowsheet incorporating pre-concenttion, heap leaching and electrolytic recovery have been calculated to be as follows:

Elecrical power	114.6×10^6 kWh/year
Heat	10.5×10^6 BTU/h

Capital Cost Estimate

The capital cost for the standard hydrometallurgical flowsheet incorporating pre-concentration, vat leaching and electrowinning of manganese metal has been estimated as follows:

Crushing and grinding	$ 6,188,000
Spiral concentration	536,000
Initial vats and ponds	486,000
Solution handling	527,000
Impurity precipitation	3,199,000
Clarification	809,000
Electrowinning	13,731,000
Anolyte treatment	7,767,000
Tailing neutralization & ammonia recovery	455,000
Reagent storage and distribution	1,985,000
Process air, vacuum and steam	539,000
Total Plant Cost	36,222,000
Mobile equipment	577,000
Contingency, 20%	7,244,000
Total Plant Capital Cost Estimate	$44,043,000

Operating Cost Estimate

The operating cost for this process alternative has been estimated as follows:

	Per Year
Reagents	$ 4,189,000
Labour	3,699,000
Electrical energy	5,730,000
Fuel oil	665,000
Operating supplies	765,000
Vat preparation	423,000
Maintenance	1,542,000
Total Operating Cost	$17,013,000
Contingency, 10% of Total Operating Cost	1,701,000
Total Plant Operating Cost Estimate	$18,714,000
Cost per kg Mn metal contained in product	$1.87

4.7 Pre-concentration and Vat Leaching for the Standard Hydrometallurgical Flowsheet with Solvent Extraction for Production of a High Purity Chemical Product

Process and Plant Summary

In this process alternative, it is suggested that solution purification by solvent extraction be used to yield a high purity chemical precipitate. This product could have the same purity (per specific listed impurities) as electrolytic manganese and thus the same market value in certain applications. The solvent extraction process would replace the iron-gypsum purification process in the standard hydrometallurgical flowsheet.

Run-of-mine ore is ground to 20 mesh in a SAG mill and concentrated in spiral concentrators. The concentrate is leached in vats and rejects are discharged to tailings. The leach solution bleed is clarified and fed to a solvent extraction circuit which extracts the dissolved manganese. The organic reagent is advanced to the stripping section where the extracted manganese is stripped off the organic. Stripped organic is recycled to extraction. A high purity manganese chemical product is precipitated from the strip solution. The barren solution is treated for reagent recovery and effluent control. The recovered reagents are recycled to the process. Details of this process alternative are presented in Appendix VII.

Design Criteria

The basic design criteria for the solvent extraction alternative are shown below. Additional criteria are presented in Appendix VII.

Ore Tonnage	- annual	401,077 tonne/year
	- average	50.64 tonne/h
	- design	55.71 tonne/h
Preconcentration		
Design	- feed	55.71 tonne/h
	- concentrate	15.31 tonne/h
	- tailing	40.40 tonne/h
	- recovery manganese	52.28%
	weight	27.48%
Production	- annual	10,000 tonne contained Mn
Product		chemical precipitate 1% listed impurities
Recovery From Concentrate		
	- heap leaching	65.0%
	- overall	63.2%
Leach Solution Bleed to Process		18.6 tonne/h
Acid consumption	- concentrate	180 kg/tonne
	- ore basis	49.5 kg/tonne ore
SO_2 consumption	- concentrate	90 kg/tonne
	- ore basis	24.7 kg/tonne ore
Lime consumption	- concentrate	128 kg/tonne
	- ore basis	35.2 kg/tonne ore

Flowsheet: Material and Energy Balance

The proposed flowsheet includes pre-concentration and vat leaching as shown in Figures No. 3 and No. 4. These circuits would replace the appropriate circuits in the standard hydrometallurgical flowsheet. The conceptual solvent extraction circuit is shown in Figure No. 5. The circuit replaces the other purification-recovery options considered in the standard hydrometallurgical flowsheet.

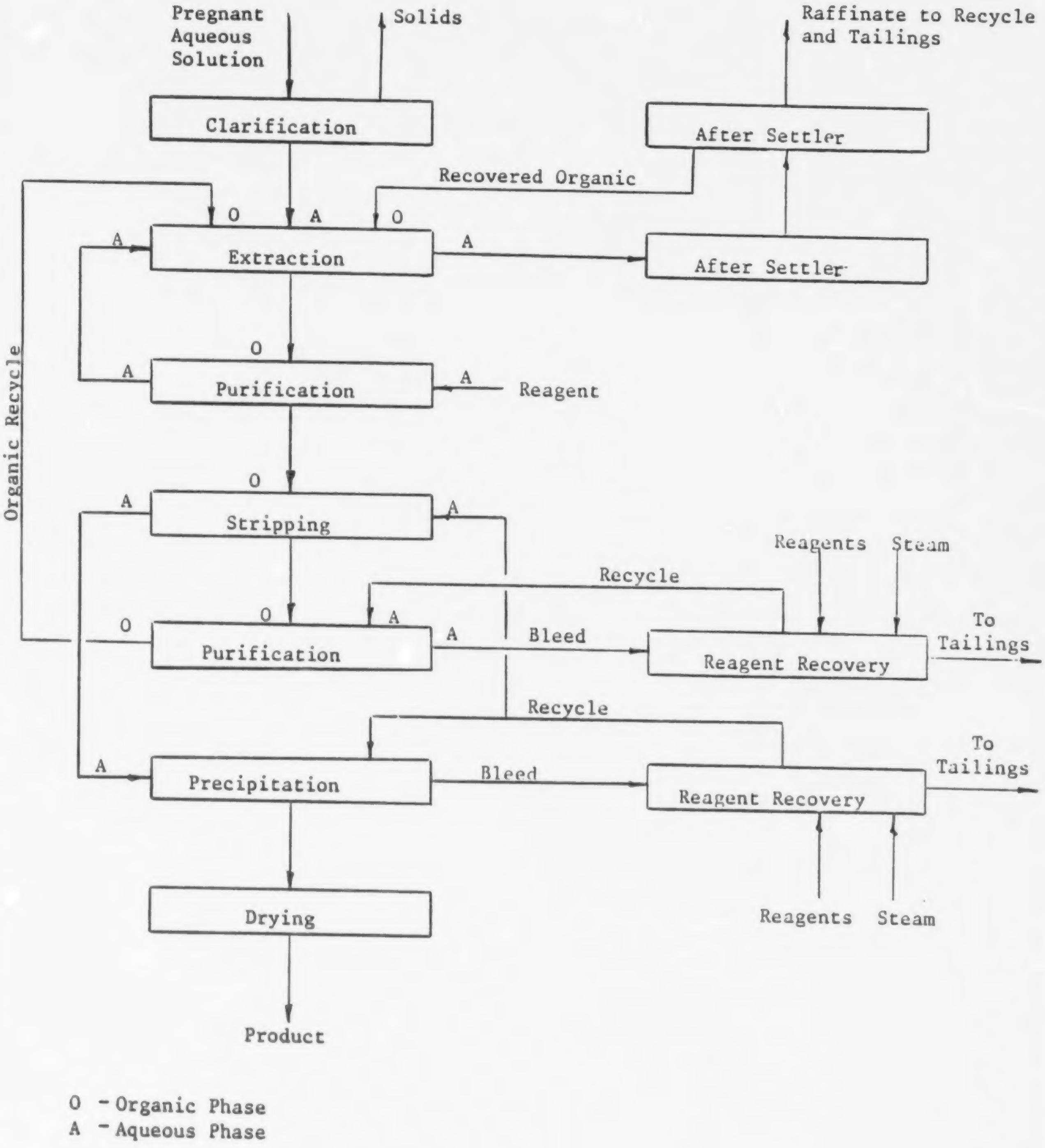
A material balance for this process alternative is presented in Appendix VII.

Energy requirements for the complete standard hydrometallurgical flowsheet incorporating pre-concentration, vat leaching and solvent extraction recovery have been calculated to be as follows:

Electrical energy	10.1×10^6 kWh/year
Heat	23.8×10^6 BTU/h

FIGURE No. 5

SOLVENT EXTRACTION PROCESS FLOWSHEET


Pregnant Aqueous Solution
Solids
Raffinate to Recycle and Tailings
Clarification
After Settler
Recovered Organic
O
A
O
A
Extraction
A
After Settler
Organic Recycle
O
A
Purification
A
Reagent
O
A
Stripping
A
Reagents
Steam
Recycle
O
O
A
Purification
A
Bleed
Reagent Recovery
To Tailings
Recycle
A
Precipitation
Bleed
Reagent Recovery
To Tailings
Reagents
Steam
Drying
Product
O - Organic Phase
A - Aqueous Phase

Capital Cost Estimate

The capital cost for this process alternative utilizing solvent extraction for recovery of a high purity chemical product has been estimated to be as follows:

Crushing and grinding	$ 6,188,000
Spiral concentration	536,000
Initial vats and ponds	486,000
Solution handling	527,000
Clarification	550,000
Solvent extraction	2,820,000
Manganese precipitation; reagent recovery	6,041,000
Tailing neutralization	203,000
Reagent storage and distribution	1,358,000
Process air, vacuum and steam	767,000
Total Plant Cost	19,476,000
Mobile equipment	544,000
Contingency, 20%	3,895,000
Total Plant Capital Cost Estimate	$23,915,000

Operating Cost Estimate

The operating cost for this process alternative has been estimated as follows:

	Per Year
Reagents	$ 9,502,000
Labour	2,153,000
Electrical energy	510,000
Fuel oil	1,470,000
Operating supplies	298,000
Vat preparation	423,000
Maintenance	837,000
Total Operating Cost	15,193,000
Contingency, 10%	1,519,000
Total Plant Operating Cost Estimate	$16,712,000
Cost per kg Mn contained in product	$1.67

4.8 Conventional Hydrometallurgical Process for $MnSO_4$

Process and Plant Summary

This process alternative produces an agricultural grade (75% $MnSO_4$ min.) manganese sulphate product which can be used for:

- Cattle feed supplement
- Fertilizer amendment

The flowsheet is the same as for the base case from crushing through to impurity precipitation. The product is recovered by crystallization rather than by chemical precipitation.

Run of mine ore is crushed and then ground in a SAG - ball circuit to 80% -200 mesh. The ground product is thickened, filtered and repulped as leach feed.

The leaching circuit is conventional as described previously with addition of steam, sulphuric acid and sulphur dioxide to maintain reducing conditions. Leach discharge is filtered in two filtration stages for recovery of pregnant solution containing the dissolved manganese. Recycle mother liquor at 8 g/L $MnSO_4$ is used for interstage repulping and wash solution.

The pregnant solution is purified by the addition of lime to pH 4.5 which precipitates impurities.

The pure solution is heated in heat exchangers by steam and recovered vapour from the crystallizer flash tank to just below the crystallizing temperature of 140°C. Final temperature elevation to 170°C is done in jacketed autoclaves using steam.

Manganese sulphate crystallizes as $MnSO_4 \cdot H_2O$. The crystallized manganese sulphate is removed in a thickener and filter circuit and dried to the anhydrous form in a multi-hearth dryer at 300°C.

A more detailed process description is presented in Appendix VIII.

Design Criteria

The basic design criteria for the conventional hydrometallurgical process to produce manganese sulphate are presented below. The design criteria necessary for engineering calculations for this process alternative are presented in Appendix VIII.

Ore Tonnage	- annual	172,000 tonne
	- average	21.7 tonne/h
	- design	23.9 tonne/h
Ore Grade		7.57 % Mn
Production	- annual	10,000 tonne contained Mn
Product		$MnSO_4$
Recovery	- leaching	80.0%
	- overall	76.8%
Pregnant Solution Flow to Purification		3,695 tonne/h
Acid Consumption		180 kg/tonne ore
SO_2 Consumption		90 kg/tonne ore
Lime Consumption		56 kg/tonne ore

Flowsheet - Material and Energy Balance

The flowsheet for this process alternative is presented as Figure No. 6. The flowsheet shows the unit operations which differ from the base case flowsheet.

The relevant material balance details are presented in Appendix VIII.

Energy requirements for the complete conventional hydrometallurgical flowsheet for production of manganese sulphate have been calculated as follows:

Electrical energy	10.6×10^6 kWh/year
Heat	30.0×10^6 BTU/h

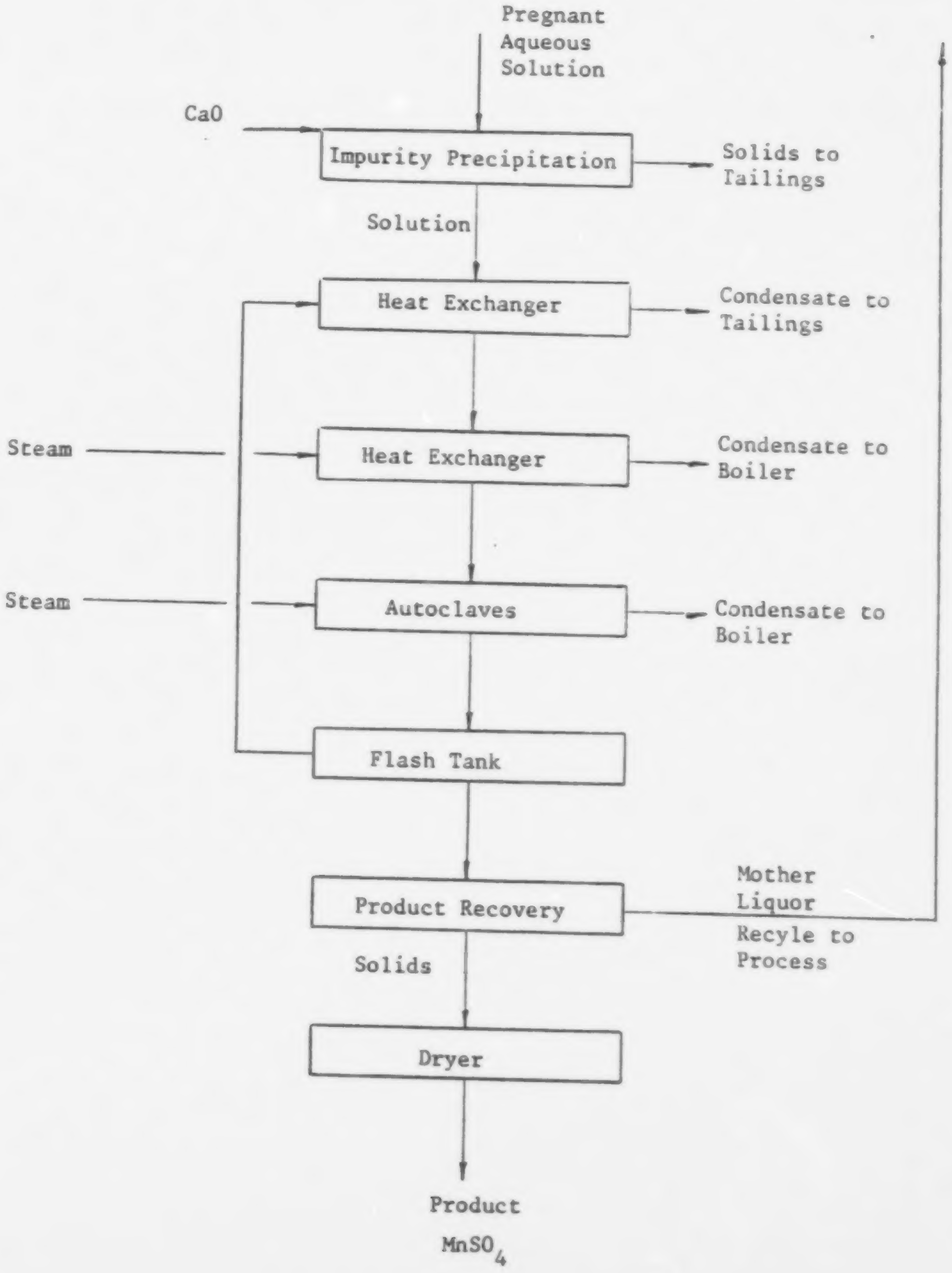
Capital Cost Estimate

The capital cost for this conventional hydrometallurgical process for recovery of manganese sulphate has been estimated as follows:

Crushing and grinding	$ 5,659,000
Neutral thickening and filtration	1,362,000
Leaching	3,042,000
Acid filtration	3,759,000
Impurity precipitation	2,461,000
Manganese sulphate crystallization and product drying	7,861,000
Tailing neutralization	452,000
Reagent storage and distribution	1,358,000
Process air, vacuum and steam	1,386,000
Total Plant Cost	$27,340,000
Mobile equipment	544,000
Contingency, 20%	5,468,000
Total Plant Capital Cost Estimate	$33,352,000

FIGURE No. 6

MANGANESE SULPHATE PROCESS FLOWSHEET


Pregnant Aqueous Solution
CaO
Impurity Precipitation
Solids to Tailings
Solution
Heat Exchanger
Condensate to Tailings
Steam
Heat Exchanger
Condensate to Boiler
Steam
Autoclaves
Condensate to Boiler
Flash Tank
Product Recovery
Mother Liquor
Recyle to Process
Solids
Dryer
Product
MnSO4

Operating Cost Estimate

The operating cost for this process alternative was estimated to be as follows:

	Per Year
Reagents	$ 8,198,000
Labour	2,228,000
Electrical energy	530,000
Fuel oil	1,845,000
Operating supplies	200,000
Maintenance	1,167,000
Total Operating Cost	$14,178,000
Contingency, 10%	1,417,000
Total Plant Operating Cost Estimate	$15,585,000
Cost per kg Mn contained in the product	$1.56

4.9 Stripped Hydrometallurgical Process for $MnSO_4$

Process and Plant Summary

This process is identical to that in Section 4.8, and as described in Appendix VIII, but with deletion of some equipment and circuits which may not be essential to produce a manganese sulphate product. Analysis of this option is for the overall project economic review - to show the cost variance associated with designing and operating a plant with "minimal" equipment.

A major deletion is the impurity precipitation circuit. This might result in the production of a product that is not readily saleable in the agricultural market as intended. Other deletions are:

- Primary crusher grizzly and oversize breaker
- Leach SO_2 scrubbing system
- Secondary acid leach filtration system
- Spare heat exchangers, autoclave, compressor, vacuum pump, and boiler.

This cut-back may adversly affect production on-stream time and plant availability.

Operating cost components (labour, reagents, energy, materials) must also be adjusted accordingly. Further details of the deletions are presented in Appendix IX.

Design Criteria

The basic design criteria for this process option are the same as for the process in Section 4.8.

Flowsheet- Material and Energy Balance

The flowsheet for this process option is the same as the flowsheet for Section 4.8 with the exception that the impurity precipitation circuit is deleted. The relevant material balance details for Section 4.8 were assumed.

Energy requirements for this process option were estimated as follows:

Electrical energy	9.4×10^6 kWh/year
Heat	30.0×10^6 BTU/h

Capital Cost Estimate

The capital cost for this alternative of a stripped hydrometallurgical process for recovery of manganese sulphate has been estimated as follows:

Crushing and grinding	$ 5,396,000
Neutral thickening and filtration	1,362,000
Leaching	2,979,000
Acid filtration	1,641,000
Impurity precipitation	-
Manganese sulphate crystallization and product drying	7,003,000
Tailing neutralization	452,000
Reagent storage and distribution	1,358,000
Process air, vacuum and steam	854,000
Total Plant Cost	$21,045,000
Mobile equipment	544,000
Contingency, 20%	4,209,000
Total Plant Capital Cost Estimate	$25,798,000

Operating Cost Estimate

The operating cost for this process alternative was estimated to be as follows:

	Per Year
Reagents	$ 8,163,000
Labour	2,067,000
Electrical energy	470,000
Fuel oil	1,845,000
Operating supplies	179,000
Maintenance	903,000
Total Operating Cost	$13,628,000
Contingency, 10%	1,363,000
Total Plant Operating Cost Estimate	$14,991,000
Cost per kg Mn contained in the product	$1.50

4.10 Pre-Concentration and Vat Leaching for the Conventional Hydrometallurgical Process for $MnSO_4$

Process and Plant Summary

Pre-concentration and vat leaching provide both capital and operating cost benefits. These circuit options have thus been included in this flowsheet in a further effort to determine if agricultural grade manganese sulphate can be produced at a cost lower than its market value.

The front portion of this process includes crushing, coarse grinding, spiral concentration, vat leaching and impurity precipitation circuits as described for Section 4.5 (Appendix V). The only difference is the recycle of mother liquor from the product crystallization circuit to the leaching vats for use as leach dilution.

The product recovery circuit includes manganese sulphate monohydrate crystallization and recovery, and product drying to $MnSO_4$. This portion of the circuit is identical to the process described in Section 4.8 (Appendix VIII). The only difference is that the tonnage of pregnant aqueous solution treated is lower.

Design Criteria

The basic design criteria for this process alternative for $MnSO_4$ production are presented below. The design criteria necessary for engineering calculations are shown in Appendix X.

Ore Tonnage	- annual	401,077 tonne/year
	- average	50.64 tonne/hour
	- design	55.71 tonne/hour
Ore Grade		7.57% Mn

Pre-Concentration

Design	- feed	55.71 tonne/hour
	- concentrate	15.31 tonne/hour
	- tailings	40.40 tonne/hour
	- recovery, manganese	52.28%
	weight	27.48%

Production	- annual	10,000tonne
Product		$MnSO_4$
Recovery from Concentrate	- leaching	65.0%
	- overall	63.0%
Leach Solution Bled to Process		18.6 tonne/h
Acid Consumption	- concentrate	180 kg/tonne
	- ore basis	49.4 kg/tonne ore
SO_2 Consumption	- concentrate	90 kg/tonne
	- ore basis	2417 kg/tonne ore
Lime Consumption	- concentrate	81 kg/tonne
	- ore basis	22.3 kg/tonne ore

Flowsheet Material and Energy Balances

The flowsheet for this process alternative is a combination of flowsheets for heap (vat) leaching, pre-concentration and manganese sulphate crystallization. The flowsheets for the above three processes are presented as Figures Nos. 2, 3, and 6.
The relevant material balance details are presented in Appendix X.

Energy requirements for this process alternative have been estimated as follows:

Electrical energy	10.6×10^6 kWh/year
Heat	30.0×10^6 BTU/h

Capital Cost Estimate

The capital cost for this process alternative which incorporates pre-concentration and vat leaching for recovery of manganese sulphate has been estimated as follows:

Crushing and grinding	$ 6,188,000
Spiral concentration	536,000
Initial vats and ponds	486,000
Solution handling	527,000
Impurity precipitation	1,904,000
Manganese sulphate crystallization and product drying	6,328,000
Tailing neutralization	167,000
Reagent storage and distribution	1,358,000
Process air, vacuum and steam	473,000
Total Plant Cost	$17,967,000
Mobile equipment	544,000
Contingency, 20%	3,593,000
Total Plant Capital Cost Estimate	$22,104,000

Operating Cost Estimate

The operating cost for this process alternative was estimated to be as follows:

	Per Year
Reagents	$ 5,471,000
Labour	1,992,000
Electrical energy	476,000
Fuel oil	1,168,000
Operating supplies	305,000
Vat preparation	423,000
Maintenance	774,000
Total Operating Cost	$10,579,000
Contingency, 10%	1,058,000
Total Plant Operating Cost Estimate	$11,637,000
Cost per kg Mn contained in the product	$1.16

5.0 ECONOMIC ANALYSIS OF PROCESS ALTERNATIVES

An economic analysis of the process alternatives studied was performed to determine the most attractive option.

The details of the evaluation are presented in Table 2, "Economic Analysis - Summary and Variances".

Several assumptions and costs, not listed elsewhere in this report, were assumed in preparing this evaluation.

The assumptions were:

Mining Costs

Mining cost, based on costs for other open pit operations, were assumed to be $1.50 per tonne mined. An ore to waste ratio of 1:1 was assumed, resulting in a total mining cost of $3.00 per tonne ore.

Capital Cost Depreciation

A capital cost depreciation of 20% of the total plant capital cost estimate was assumed.

Interest

Interest was assumed to be 10%.

Table 2

ECONOMIC ANALYSIS - SUMMARY AND VARIANCES

Process Flowsheet Alternative	Capital Cost $ (1000's)	Capital Cost Depreciation + Interest $/kg Mn	Operating Cost $/Year (1000's)	Operating Cost $/kg Mn	Mining Cost $/kg Mn	Total Product Cost $kg Mn	Product Value $/kg Mn	Operating Margin $/kg Mn	Capital Cost Variance from Base Case, %	Operating Cost Variance from Base Case, %	Change in Operating Cost to Breakeven %
1. Standard Hydrometallurgical Production of a Middling Chemical Product $Mn(OH)_2$	33,871	1.02	18,118	1.81	0.05	2.88	2.00	-0.88	Base	Base	-31
2. Effects of Pre-Concentration on the Standard Hydrometallurgical Process	29,232	0.88	12,946	1.30	0.10	2.28	2.00	-0.28	-14	-28	-12
3. Effects of Heap Leaching on the Standard Hydrometallurgical Process	22,116	0.66	19,184	1.92	0.06	2.64	2.00	-0.64	-35	+ 6	-24
4. Effects of Electrowinning on the Standard Hydrometallurgical Process	52,871	1.59	23,823	2.38	0.05	4.02	3.85	-0.17	+56	+32	- 4
5. Pre-concentration and Vat Leaching for the Standard Hydrometallurgical Process for $Mn(OH)_2$	22,828	0.69	13,960	1.40	0.12	2.21	2.00	-0.21	-33	-23	-10
6. Pre-concentration and Vat Leaching for the Standard Hydrometallurgical Process with Electrowinning	44,043	1.32	18,714	1.87	0.12	3.31	3.85	+0.54	+30	+ 3	+16
7. Pre-concentration and Vat Leaching for the Standard Hydrometallurgical Process with Solvent Extraction for Production of a High Purity Product	23,915	0.72	16,712	1.67	0.12	2.51	± 3.85	+1.34	-29	+ 8	+53
8. Conventional Hydrometallurgical Process for $MnSO_2$	33,352	1.00	15,585	1.56	0.05	2.61	1.06	-1.55	- 2	-14	-59
9. Stripped Hydrometallurgical Process for $MnSO_4$	25,798	0.77	14,991	1.50	0.05	2.32	1.06	-1.26	-24	-17	-38
10. Pre-concentration and Vat Leaching for the Conventional Hydrometallurgical Process for $MnSO_4$	22,104	0.66	11,637	1.16	0.12	1.94	1.06	-0.88	-35	-36	-45

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Product Value

Product value, for $MnSO_4$ and electrolytic manganese metal was calculated from published reports. The calculated costs were:

$MnSO_4$	\$1.06/kg Mn contained
Mn metal	\$3.85/kg metal

The value for a middling chemical product was not found in any publication. A value of less than \$2.00/kg Mn contained was assumed.

The following comments can be made about the alternatives evaluated.

Alternative 1 (Base Case)

Production of a middling chemical product is not economic. Direct operating costs amount to 90% of the assumed product value. Reagent costs alone are 50% of the product value and leaching reagent costs are 36% of product value. Total production costs must be reduced by 30% for a break even situation. Such a reduction in cost is unlikely.

Alternative 2 (Pre-concentration)

The inclusion of a pre-concentration circuit into the base case flowsheet results in only a slight reduction in capital cost (14%), but a significant reduction in operating cost (28%). The major portion of the operating reduction is due to a significant reduction in leaching reagent costs (48%). However, the process is not economic. Total production costs must be reduced by 12% for a break even situation. Detailed capital and operating cost estimates, based on detailed testwork might show this reduction to be possible. However, it would still be a break even situation that is completely dependent on the assumed product value.

Alternative 3 (Heap Leaching)

The heap leaching alternative for the base case significantly decreases the capital cost (35%). However, there is a slight increase in operating costs (6%), due mainly to an increase in leaching reagent requirements because of the lower leach extraction for heap leaching. Total production costs must be reduced by 25% for a break even situation. The process is not economic.

Alternative 4 (Electrowinning)

Both the capital cost and the operating costs are significantly greater for the process alternative which includes electrowinning of manganese metal (56% and 32% respectively). Estimated total production costs are slightly higher than the product value (5%) and an equivalent cost reduction would result in a break even situation. The electrowinning of manganese metal from an alkaline catholyte with an acidic anolyte is extremely difficult. Therefore, it is not expected that detailed cost estimates based on detailed testwork will show an improvement in total production cost. In fact, the current estimates might be low. The process is not economic.

Alternative 5 (Combination of Pre-Concentration and Heap Leaching)

The inclusion of pre-concentration and heap leaching into the base case flowsheet preserves almost all the capital and operating cost advantages of each of the individual inclusions. However, the process is still not economic, and total production costs must be reduced by 10% to achieve a break even situation.

Alternative 6 (Combination of Pre-concentration, Heap Leaching and Electrowinning)

A combination of the above three process options still results in a significantly higher capital cost than the base case (30%), but results in almost equal operating costs. The operating cost reduction for this option is due mainly to savings in leaching reagents (acid). The solution balance for the process indicated that total recycle of spent anolyte to leaching might be possible. This results in use of the acid generated in the cells and only minor quantities of fresh acid addition are necessary. The process is indicated to be marginally economic. However, an increase in total cost of 14% would result in a break even situation. In light of the complexity and difficulty of the electrowinning of manganese metal, such an increase would not be unexpected.

Alternatives 8, 9 and 10 ($MnSO_4$ Alternatives)

All three $MnSO_4$ production options are not economic irrespective of improvements in capital and operating cost because of process innovations. In each case, the direct operating cost is greater than the product value. Again leaching reagents are the main contributors to the operating costs; making up 82, 81 and 55% of the costs respectively. The low product value makes these processes uneconomic. The capital costs for the conventional $MnSO_4$ alternative and the combined pre-concentration heap leach alternatives are similar to the corresponding two alternatives for the base case. Operating costs are lower (14% and 36% respectively). The costs for the stripped circuit did not improve the operating margin. Total cost reductions in the order of 33-59% are required for a break even situation. At this time, reductions of this magnitude are considered unrealistic.

Alternative 7 (Combination of Pre-concentration, Heap Leaching and Solvent Extraction)

The indicated operating margin for this process alternative for the base case is 53%. Total operating costs can increase by this percentage before a break even situation is reached.

The capital cost for this option is 30% lower than the base case, but well within the range of capital costs estimated for several other options. The operating cost is only 8% lower than base case. A most significant factor is the expected high value of the high purity product. At this time, there is no reason to expect a lower value for the chemical product than for the equivalent purity electrolytic metal. The proposed product could be used as a substitute for the metal, in applications where the metal must be converted to the desired chemical. If a suitable solvent extraction process can be developed, total operating costs are expected to be well within the 53% margin available.

6.0 CONCLUSIONS

A review of the economic analysis shows the following.

The direct operating costs for the manganese sulphate ($MnSO_4$) products are greater than the apparent value of the product. The major cost factor is the reagent cost. However, even with significant reductions in reagent cost, total production costs are likely to remain greater than the product value. The New Brunswick manganese ore is not well suited to the production of low value products.

In most cases evaluated, the high operating costs are directly influenced by the high cost of the leaching reagents, sulphuric acid and sulphur dioxide. Several possibilities for reducing these reagent costs exist:

- negotiation of lower prices for long term contracts
- recovery from metallurgical off-gases available regionally
- preparation from pyrite tailings available locally
- preparation from purchased sulphur

Excluding lower negotiated prices, the other options for reagent cost reduction would require significant capital expenditures with the net result that total production costs might remain relatively high.

The reagent prices listed were reviewed in depth with suppliers. The prices quoted appeared to be reasonable, but there also appeared to be some room for reductions. Reductions in the order of ten per cent might be achievable with firm contract commitments.

The high capital and operating costs for the plants make the production of a middling chemical product uneconomic. The high capital costs are a function of the process. The handling and processing of acidic pulps and solutions almost doubles process plant costs in comparison with neutral or basic operations such as cyanidation for gold recovery. Acid resistant equipment must be used for acidic circuits and this equipment with the associated process piping and other plant features is expensive and there is no means of reducing costs.

The value of the middling chemical product has been assumed to be $2.00/kg contained manganese. However, the middling product could be considered only suitable as a feed material for further purification such as electrowinning. Under these circumstances the value of the product might be in the order of $1.00/kg contained manganese. This is the estimated potential savings in terms of capital and operating cost that might be realized in using the middling product in place of the New Brunswick ore for production of a high purity product.
On this basis, comparisons with ores that are less expensive to treat than the New Brunswick ore might result in a lowering of the value of the middling product even further.

Gravity pre-concentration of the ore prior to hydrometallurgical processing significantly reduces the operating costs by reducing leaching reagent requirements. The pre-concentration process appears to reject acid consumers. The capital cost is also decreased with this ore pre-treatment. Pre-concentration should be included in the flowsheets for manganese recovery from New Brunswick ore.

Heap (or vat) leaching significantly reduces capital cost requirements. However, operating costs, mainly leaching reagent cost, increase significantly because a greater ore tonnage must be treated to achieve production goals. A combination of pre-concentration and heap leaching (heap leaching of gravity concentrate) provides the advantages of both options in terms of lower operating and lower capital cost respectively. This combination of these process options should be included in the flowsheets for manganese recovery from New Brunswick ores.

Electrowinning of manganese metal using a standard hydrometallurgical flowsheet is not economic. The alternative is marginally economic when pre-concentration and vat leaching are included in the flowsheet. A major cost component is the electrical energy cost.

The electrowinning of manganese depends on very pure solutions. The process depends on a combination of factors such as concentration, pH, temperature, current density, cathode and anode material, rate of flow to the cells, as well as other variables. Problems of dealing with an alkaline catholyte and an acid anolyte in a two-compartment cell, as well as the impossibility of considering any of the variables separately because of their influence on each other, cause manganese electrowinning to be almost in a class by itself. It demands much more delicate control than does the electrowinning of zinc and copper, for it carries all their problems and many more.

The slight operating margin indicated is not considered adequate to encourage pursuit of a complicated process such as electrowinning of manganese. At this time, it is concluded that the electrowinning process alternative is not applicable to the New Brunswick ore.

The process alternative with pre-concentration, heap leaching and solvent extraction provides the best operating margin. Total production costs could increase by up to 53% to provide a break even situation. The cost data indicate that a standard hydrometallurgical process (not including pre-concentration or heap leaching) with solvent extraction purification and recovery would still be economic. The capital and operating costs for such a process are estimated at $35 million and $2.10/kg contained manganese respectively. The resultant total operating cost would be approximately $3.20/kg of contained manganese, providing an operating margin of $0.65 /kg. This margin provides for an additional 20% increase in total production costs for a break even situation.

[1]Work of a preliminary nature has shown that solvent extraction for manganese purification and recovery is technically feasible. In addition, Witteck Development Inc. has developed economic solvent extraction processes for recovery of a high purity product of another metal for which solvent extraction was not initially considered At this time, the probability of developing an economic solvent extraction process for manganese, where technical feasibility has already been shown, is considered acceptable. This preliminary economic evaluation of the process provides the necessary incentive to investigate this alternative.

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1. G.M. Ritcey and B.H. Lucas. Purification of manganese solutions containing copper and zinc by liquid-liquid extraction, using di-(2-ethylhexyl) phosphoric acid. Canadian Metallurgical Quarterly, Volume 10 Number 3 (1971).

The economic evaluation clearly shows that conventional hydrometallurgical processes are not viable for the New Brunswick ore at the production rate of 10,000 tpy contained manganese. This conclusion will remain valid unless key operating cost components can be decreased and higher tonnage rates be supported by the product market.

It is suggested that further work be directed to the development of new technology for the production of high value products, for without this change in strategy, the future of the project is considered marginal at best.

7.0 RECOMMENDATIONS

On the basis of the economic analysis and the conclusions, it is recommended that a research program be initiated to develop a hydrometallurgical process with solvent extraction for purification and recovery of a high purity manganese product from the New Brunswick ore.

The program can proceed in stages and a suggested approach is as follows:

- Literature review of possible solvent extraction processes.
- Batch scoping tests of perceived viable alternatives.
- Batch test confirmation of the technical feasibility of the selected process(es).
- Economic evaluation of the selected process(es) in terms of capital and operating costs including auxiliary processes such as product recovery, reagent recovery and recycle, and effluent control.
- Confirmation of the practical feasibility of the selected process(es) by operation of a closed loop bench scale mini-pilot plant. Information developed would include data on impurity build-up, solution balances, reagent consumptions, process efficiencies, product specifications, and process operability.

- Re-evaluation of the process economics on the basis of the mini-pilot plant data.

- Operation of a pilot plant of ample size to provide detailed information on process design, performance optimization, equipment design factors, scale up factors, reagent requirements, effluent quality and product quality. Suitable samples of the product(s) would be prepared for evaluation by potential customers.

- Detailed economic evaluation of the overall project economics.

The stage approach permits suggested periodic progress review in terms of economic evaluation. If the evaluations show that the economics are unfavourable, the program can be terminated or revised, reducing non-productive expenditures.

Work to date has shown that pre-concentration and heap (vat) leaching provide significant reduction in operating and capital costs. These two process alternatives should be studied in detail to confirm the economics and to optimize performance. These process alternatives should be considered for the mini-pilot plant program, but must be included in the pilot plant program.

Economic studies should include evaluation of the options for supplying leaching reagents. Reductions in reagent costs will improve the economics of all process alternatives.

The feasibility of this project, as all other projects, is contingent on a market for the product. A market survey is necessary to determine the marketability of a high purity chemical product.

APPENDIX I

STANDARD HYDROMETALLURGICAL PROCESS FOR PRODUCTION OF A MIDDLING CHEMICAL PRODUCT, $Mn(OH)_2$

PROCESS AND PLANT DESCRIPTION

From the run-of-mine stockpile, ore is conveyed by front end loader to the grizzly. Large ore is broken with a rock breaker and all ore is passed through a jaw crusher. The crusher operates 2 shifts per day, 5 days per week and feeds both the milling circuit and a crushed ore stockpile. During crusher down time ore is retrieved from the crushed ore stockpile and introduced into the milling circuit.

The grinding circuit consists of a SAG and ball mill in closed circuit with cyclones. Ore is ground to a nominal 80% -200 mesh. Final product, the secondary cyclone overflow, is directed to the neutral thickener. After thickening and dewatering on a disc filter, the wet cake is reslurried with a portion of the pregnant solution. The resulting slurry is pumped directly to the leach circuit.

Leaching of the manganese ore is accomplished in a cascading system of leaching vessels. The tanks are rubber lined to withstand the acidic environment. Sulphuric acid is added to the lead reactor at the rate of 180 kilograms per tonne. Leach density is maintained at approximately 50% solids.

During the 24 hour leach period, 15% of the solids are dissolved and 80% of the contained manganese is extracted. Sulphur dioxide is introduced into the leaching vessels to maintain reducing conditions. The reducing conditions are necessary to solubilize manganese and minimize the dissolution of iron. Unused and escaped gas is scrubbed in a packed column with a counter-flow of lime.

To improve extraction kinetics, leaching is carried out at 80°C by injection of recovered vapours from the evaporators in the reagent recovery circuit. Leach discharge is directed to the acid filtration circuit for recovery of solubilized manganese.

Two stages of drum filters process the solids discharging the leach circuit to minimize soluble losses. Drum sizing is based on a filterability of 270 kilograms per square meter per hour. Each filtration stage is comprised of 4 filter units.

Leach discharge goes directly to a head tank and is distributed to the boots of the primary drum filters. The filtrate drawn from the secondary filters is used to wash the primary filter cake. Primary drum filtrate is collected in vacuum system tankage and distributed to the neutral disc filter repulper and the iron-gypsum precipitation circuit. Wet primary filter cake is conveyed to a mixing tank where it is reslurried with process water. This slurry is distributed to the secondary drum filters. On the secondary drum filters the cake is washed with process water and the moist cake conveyed to the tailings neutralization circuit.

A portion of the leach pregnant solution is used for dilution of leach feed. The balance of the pregnant solution is pumped to the head of the iron-gypsum precipitation circuit.

The iron-gypsum cascade system of 4 stirred tank reactors has the function of bleeding sulphates from the system as gypsum and removing impurites such as iron and aluminum. Removal of the sulphate is accomplished by addition of calcium chloride which converts the sulphate solution to a chloride solution. Removal of the sulphates permit the later precipitation of manganese with lime. Contamination of the final product by calcium sulphate is reduced.

Chloride conversion is completed in the first two stages of the circuit. Lime slurry is introduced into the third stage to raise the pH to a suitable level. At this point iron is precipitated from the solution and residual sulphuric acid is removed as gypsum.

After thickening and filtering, a moderately clean manganese chloride solution is available for manganese recovery. The precipitate after washing is conveyed to the tailings neutralization circuit.

The purified manganese chloride solution from the iron-gypsum circuit is pumped to the head of the manganese precipitation circuit. This circuit consists of a cascade arrangement of two stirred tank reactors. Lime is added to raise the pH to 9.5. Manganese precipitates as the hydroxide and chloride ion reverts to calcium chloride.

The slurry from the manganese precipitation circuit is pumped to a thickener, thickened, filtered and washed on a drum filter. The washed cake discharge is conveyed into a directly fired multiple hearth furnace. Dry product is discharged into a silo for storage and loading into tank trucks.

Because it contains calcium chloride, effluent from the manganese recovery circuit must be recycled. The recycle is introduced at the head of the iron-gypsum circuit. To maintain a solution balance some of the effluent must be evaporated off. Product thickener overflow is pumped to a four effect falling film evaporator. The concentrated calcium chloride solution is recycled to process. The vapours from the last evaporator are used to heat the leach tanks and for heating the process building.

Washed cake from the secondary acid filters, washed iron-gypsum precipitate and leach off-gas scrubber effluent are fed to the tailings neutralization circuit. The final tailings slurry pH is raised to 10.5 by addition of lime slurry. Recycle tailings decant is used for pulp dilution.

The tailings neutralization circuit discharge is pumped directly to the tailings management area. No consideration is given here to the nature of the tailings management area or to its operation and cost.

The reagent storage area includes facilities for receipt, storage, preparation and distribution of the five major reagents.

Sulphuric acid arrives in railcars possibly from the Sudbury area. The cars are off-loaded into one of two steel storage tanks. Centrifugal pumps advance the process requirements to the leach tank circuit.

Sulphur dioxide also arrives from the Sudbury area in railcars. From the railcar it is transferred to the storage tank. Process requirements are pumped from storage and passed through a vapourizer unit.

Calcium Chloride is shipped in 22 kilogram sacks on non-returnable pallets. The sacks are broken open and their contents placed is a storage silo. The silo is located adjacent to the lead tank of the iron-gypsum circuit. An adjustable screw discharge introduces the required amount of dry crystals directly into the pregnant solution.

Lime is available locally and is delivered by tanker truck. Lime is discharged pneumatically into a storage silo. A bin activator and screw conveyor deliver lime directly to a lime slaking unit which overflows to a day tank. Water for slacking is drawn from the process water circuit. The 25% solids slurry is pumped continuously through a lime distribution circuit and drawn off at the three demand points.

It is assumed that a non-ionic type flocculant will be suitable for this process. The product will be delivered to the site in sacks. A 1% flocculant solution is prepared in the mix tank and then transferred to a day tank where water is added to dilute the mixture to a 0.1% solution. From the day tank, the flocculant is circulated continuously through a distribution system and drawn off at the demand points.

Plant services such as compressed air and steam are provided by a powerhouse area which also includes the process air compressors, the process vacuum pumps and the process steam boilers. As is normal practice, the process services distribution system is separate from the plant service system.

DESIGN CRITERIA

The basic design criteria for this hydrometallurgical process, the base case are:

Ore Tonnage	-annual	172,000 tonne
	-average	21.7 tonne/h
	-design	23.9 tonne/h
Ore Grade		7.57% Mn
Production	-annual	10,000 tonne contained Mn
Product		$Mn(OH)_2$, 60% Mn
Recovery	-leaching	80.0%
	-overall	76.8%
Pregnant Solution Flow to Purification		36.95 tonne/h
Acid Consumption		180 kg/tonne ore
SO_2 Consumption		90 kg/tonne ore
Lime Consumption		115 kg/tonne ore
$CaCl_2$ Consumption		10.7 kg/tonne ore

The engineering design criteria for equipment sizing, and the material balance are as follows:

Crushing and Grinding:

Crushing	-schedule	2 shifts per day 5 days per week
	-rate	50.2 tonne/h
Grinding	-schedule	24 hours per day 7 days per week
	-rate	23.9 tonne/h
Grinding Mills		primary SAG secondary Ball
Grind		80% minus 200 mesh
Grinding Energy		15.5 kWh/tonne
Classification		cyclones

Neutral Thickening and Filtration

Thickener	-unit area	0.30m^2 /tonne/day
	-underflow density	50% solids
	-flocculant	0.125 kg/tonne
Filtration	-rate	270 kg/m^2/h
	-cake moisture	20%
Filter	-type	disc
	-number	1

Leaching

Leach	-density	40-60% solids
	-temperature	78°C
	-retention time	24 h
	-pH	2.0
	-extraction	80%
Leach Tanks	-number	6
	-type	agitated tank - gas reactors
Heating		evaporator discharge direct injection
SO_2	-addition	90 kg/tonne
	-loss	10%

Acid Filtration

Filtration	-rate	270 $kg/m^2/h$
	-cake moisture	20%
	-wash ratio	1:1
Wash	-primary	secondary filtrate
	-secondary	process water
Filter	-type	drum
	-number	4 per stage
Pregnant Solution Flow to Purification		36.95 tonne/h

Iron-Gypsum Precipitation

Gypsum Precipitation	-reagent	$CaCl_2$
	-retention time	2 h
Gypsum Tanks	-type	agitated tanks
	-number	2
Iron Precipitation	-reagent	CaO
	-retention time	2 h
	-pH	4.5 - 6.0
Iron Tank	-type	agitated vessels
	-number	2
Thickening	-size basis	0.2 m/h upflow
	-underflow density	20% solids
	-flocculant	as necessary
Filtration	-rate	270 $kg/m^2/h$
	-cake moisture	50%
	-wash ratio	1:1
	-wash	process water
Filter	-type	drum
	-number	2

Manganese Precipitation and Reagent Recovery

Manganese Precipitation	-reagent	CaO
	-pH	9.5
	-retention time	2 h
Precipitation Tanks	-type	agitated tanks
	-number	2
Thickening	-size basis	0.2 m/h upflow
	-underflow density	20% solids
	-flocculant	as necessary
Filtration	-rate	270 kg/m^2/h
	-cake moisture	50%
	-wash ratio	1:1
	-wash	process water
Filter	-type	drum
	-number	1
Drying	-product moisture	1%
Dryer	-type	multiple hearth
	-fuel	oil
	-temperture	300°C
Product		$Mn(OH)_2$
Overall Plant Recovery		76.8%
Reagent Recovery	-reagent	$CaCl_2$
	-overall recovery	95%
Evaporation	-type	falling film
	-number	4 stages
	-heat requirement	400 BTU/lb water
	-heat source	steam

Tailings Neutralization

Feed		all solid and solution discharges
Neutralization	-reagent	CaO
	-pH	10.5
	-retention time	3 hours
	-discharge density	35% solid
Tanks	-type	agitated vessels
	-number	3

Reagent Storage and Distribution

Sulphuric Acid Storage Tank	-reagent	H_2SO_4, 93%
	-capacity	2 week consumption
	-number	1
Sulphuric Dioxide Storage Tanks	-reagent	SO_2, liquid
	-capacity	1 week each
	-number	2
Evaporator	-number	1
Calcium Oxide Storage Bin	-reagent (Lime)	CaO
	-capacity	2 week consumption
	-number	1
Lime Slaking	-type	Dorr
	-number	1

Calcium Chloride	-reagent	$CaCl_2$
Storage	-capacity	truckload
	-type	bags
Flocculant	-reagent	polyacrylomide
Storage	-capacity	truck load
	-type	bags
Make-up Tanks	-type	agitated vessels
	-number	2
Fuel oil	-heat value	160,000 BTU/US gallon
Storage Tanks		2

Process Air, Vacuum and Steam

Air requirements		0.3 CFM/ft^2 filter area
Vacuum requirements		2 CFM/ft^2 filter area
Boilers	-tpye	oil
	-efficiency	80%

MATERIAL BALANCE DETAILS

The material balance for this process alternative is presented on the flowsheet, Drawing Number 100-110-001 Revision P1. The volumetric flowrates are approximate, and are suitable for sizing process equipment.

The vapor discharge from the fourth evaporator stage is used for heating leach pulp and to heat the building. The resultant condensate from the building heaters is discharged to tailings.

CAPTIAL COST DETAILS

Crushing and Grinding

Equipment	No.	Size	Operating Power, kW	Unit Cost, $	Total Cost, $
Grizzly	1	610 mm	-	11,000	11,000
Coarse ore bin	1	3.6 m x 3.6 m x 6.8 m	-	21,000	21,000
Hydraulic rock breaker	1	4.88 m	22.6	64,000	64,000
Coarse ore feeder	1	990 mm x 4.27 m	18.6	40,000	40,000
Jaw crusher	1	510 mm x 203 mm	111.9	124,000	124,000
Belt conveyor	3	762 mm x 30 m	3x34.8	53,333	160,000
Dust collector c/w fan	1	11,500 m^3/h	11.3	19,000	19,000
Belt scale	1	762 mm wide	0.2	8,000	8,000
Sump pump	1		3.8	8,000	8,000
SAG mill	1	3.81 m x 1.52 m	263	758,000	758,000
Ball mill	1	2.13 m x 3.05 m	183	294,000	294,000
Pump box	1	1.6 m x 1.6 m x 1.6 m	-	10,000	10,000
Primary cyclone feed pumps	2	-	15	5,000	10,000
Primary cyclone	4	-	-	8,250	33,000
Secondary cyclone	6	-	-	6,666	40,000
Pump box	1	1.6 m x 1.6 m x 1.6 m	-	10,000	10,000
Secondary cyclone feed pump	2		38	3,500	7,000
Total			771.8		$1,617,000

Total new equipment cost	$1,617,000
Equipment installation	440,000
Piping, materials and installation	440,000
Electrical, materials and installation	440,000
Instrumentation, materials and installation	210,000
Main mill building	878,000
Auxiliary building allowance	263,000
Plant services, water, air, steam etc.	263,000
Site preparation	317,000
Field expenses	210,000
Project engineering, management, accounting, purchasing	581,000
Total Area Cost	$5,659,000

Total Area Cost = Total new equipment cost x 3.50
= $1,617,000 x 3.50 = $5,659,000

Neutral Thickening and Filtration

Equipment	No.	Size	Operating Power, kW	Unit Cost, $	Total Cost, $
Neutral thickener	1	14.7 m x 1.22 m	2	142,000	142,000
Thickener O/F pump box	1	1 m x 1.22 m	-	1,000	1,000
O/F pump	2	-	2	3,600	3,600
Process water tank	1	7 m x 7 m	-	67,000	67,000
Process water pump	2	-	3.4	3,400	6,800
U/F pump	2	-	12	4,200	8,400
Leach feed surge tank c/w agitator	1	2.3 m x 2.8 m	8.8	16,000	16,000
Leach feed pump	2	-	12	4,200	8,400
Disc filter	1	2.75 m x 7 discs	3.75	131,000	131,000
Filtrate receiver	1	0.6 m x 1.9 m	-	1,000	1,000
Moisture trap	1	0.75 m x 1.9 m	-	1,200	1,200
Seal tank	1	-		1,000	1,000
Filtrate pumps	2	-	1.5	1,000	2,000
Total			45.45		389,400 (389,000)

Total Area Cost = $389,000 x 3.50 = $1,362,000

Leaching

Equipment	No.	Size	Operating Power, kW	Unit Cost, $	Total Cost, $
Leaching tanks c/w agitators	6	5 m dia.x 5.5 m	6 x 18.8	119,000	714,000
Filter feed surge tank c/w agitator	1	5 m dia.x 5.5 m	18.8	119,000	119,000
Filter feed pump	2		5.6	18,000	18,000
SO_2 scrubber c/w fan	1	-	11.25	14,000	14,000
Scrubber pump	2	-	2.3	4,000	4,000
Total			150.75		869,000

Total Area Cost =$869,000 x 3.50 = $3,042,000

Acid Filtration

Equipment	No.	Size	Operating Power,kW	Unit Cost,$	Total Cost,$
Primary drum filters	4	3.1 m dia.x 3.1 m	3 x 2	72,000	288,000
Discharge conveyor	1	914 mm x 13.7 m	7.5	50,000	50,000
Repulper c/w agitator	1	2 m dia. x 2 m	5.6	16,000	16,000
Pump box	1	2 m dia. x 2 m		7,000	7,000
Discharge pump	2		5.6	9,000	18,000
Filtrate receiver	2	0.6 m x 1.2 m		4,000	4,000
Filtrate pump	2		5.6	3,700	7,400
Moisture trap	1	0.75 m x 1.9 m		2,500	2,500
Seal tank	1			1,000	1,000
Filtrate surge tank	1	5 m dia. x 5.5 m		67,000	67,000
Filtrate pump	2		5.6	3,700	7,400
Secondary filter feed surge c/w agitator	1	5 m dia. x 5.5 m	18.8	119,000	119,000
Secondary filter feed pump	2	-	5.6	9,000	18,000
Secondary drum filter	4	3.1 m dia.x3.1 m	3x2	72,000	288,000
Discharge conveyor	1	914 mm x 13.7 m	7.5	50,000	50,000
Repulp tank c/w agitator	1	2 m dia. x 2 m	5.6	16,000	16,000
Pump box	1	2 m dia. x 2 m		7,000	7,000
Discharge pump	2	-	5.6	9,000	18,000
Filtrate receiver	2	0.6 m x 1.9 m		4,000	4,000
Filtrate pump	2	-	5.6	3,700	7,400
Moisture trap		0.75 m dia.x1.9 m	2,500	2,500	2,500
Filtrate surge tank	1	5 m dia. x 5.5 m		67,000	67,000
Seal tank				1,000	1,000
Filtrate pump	2	-	5.6	3,700	7,400
Total			96.2		$1,073,600 ($1,074,000)

Total Area Cost = $1,074,000 x 3.50 = $3,759,000

Iron-Gypsum Precipitation

Equipment	No.	Size	Operating Power, kW	Unit Cost, $	Total Cost, $
Iron-Gypsum Precipitation tanks c/w agitator	4	4.5 m dia.x5.0 m	4x11.25	54,000	216,000
Discharge pump	2	-	11.25	15,000	30,000
Iron-Gypsum thickener	1	22 m dia.	3.75	339,000	339,000
Filter feed tank c/w agitatior	1	5 m dia x 5.5 m	18.8	119,000	119,000
Drum Filter	2	3.1 m dia.x3.1 m	2x2	72,000	144,000
Filtrate receiver	1	0.6 m x 1.9 m		2,000	2,000
Mositure trap	1	0.6 m x 1.5 m		2,500	2,500
Filtrate pump	2	-	5.6	3,700	7,400
Seal Tank	1	-		1,000	1,000
Thickener U/F pump	2	-	11.25	15,000	30,000
Repulp tank c/w agitator	1	2 m dia. x 2 m	5.6	16,000	16,000
Pump box	1	2 m dia. x 2 m		7,000	7,000
Total			105.25		$913,900 ($914,000)

Total Area Cost = $914,000 x 3.50 = $3,199,000

Manganese Precipitation and Reagent Recovery

Equipment	No.	Size	Operating Power, kW	Unit Cost, $	Total Cost, $
Manganese precipitation tank c/w agitator	2	4.5 m dia x 5.0 m	2x11.25	54,000	108,000
Pump box	1	2 m dia. x 2 m		7,000	7,000
Thickener feed pump	2	-	1x5.6	9,000	18,000
Thickener	1	22 m dia.	3.75	339,000	339,000
Thickener U/F pump	2	-	1x11.25	15,000	30,000
Drum filter	1	10 m^2	1x0.25	70,000	70,000
Filtrate receiver	1	-		2,000	2,000
Moisture trap	1	-		2,500	2,500
Filtrate pump	2	-	1x5.6	3,700	7,400
Seal tank	1	-		1,000	1,000
Dryer	1	-	10	647,500	647,500
Dust collector	1	-	3.0	15,000	15,000
Belt conveyor	1	914 mm x 15 m	7.5	50,000	50,000
Product holding silo	1	5 m dia. x 7 m		37,000	37,000
Evaporator	1	4 effect		750,000	750,000
Solution recycle pump	2	-	1x3.75	3,500	7,000
Total			69.95		2,091,400

Total Area Cost = $2,091,000 x 3.50 = $7,318,000

Tailings Neutralization

Equipment	No.	Size	Operating Power, kw	Unit Cost,$	Total Cost,$
Neutralization tank c/w agitator	3	4 m dia. x 4.5	3x7.5	35,000	105,000
Pump box	1	2.0 m dia. x 2 m		3,300	3,300
Tailings pump	2	-	7.5	10,400	20,800
Total			30.0		129,100
					(129,000)

Total Area Cost = $129,000 x 3.50 = $452,000

Reagent Storage and Distribution

Equipment	No.	Size	Operating Power, kW	Unit Cost,$	Total Cost,$
H_2So_4 storage tank	1	9 m dia. x 9 m		85,000	85,000
Acid supply pump	2		3.75	3,800	7,600
$CaCl_2$ storage tank c/w screen	1	2 m dia. x 2 m	0.75	8,500	8,500
SO_2 storage tank	2	8 m dia. x 9 m		72,000	144,000
SO_2 evaporator	1			10,000	10,000
SO_2 supply pumps	2		3.75	3,800	7,600
CaO storage tank	1	7 m dia. x 8 m		16,500	16,500
Lime slaker	1		3.75	10,000	10,000
Lime distribution pump	2		3.75	3,600	7,200
Flocculant preparation tank	2	2 m dia. x 2 m	2x3.75	8,500	17,000
Flocculant transfer and distribution pump	2		1x5.2	3,400	6,800
Fuel oil storage tank	2	6 m dia. x 8 m		32,000	64,000
Fuel feed pump	2			2,000	4,000
Total			24.7		388,200 (388,000)

Total Area Cost = $388,000 x 3.50 = **$1,358,000**

Process Air, Vacuum and Steam

Equipment	No.	Size	Operating Power, kW	Unit Cost,$	Total Cost,$
Vacuum pump	3	3500 CFM	94	27,000	81,000
Compressor	2	1000 CFM	75	60,000	120,000
Boilers, complete	3	500 BHP	20	-	263,000
Total			189		$464,000

Total Area Cost = $464,000 x 3.50 =	$1,624,000

SUMMATION

Total Plant Cost	$27,772,000
Front end loader to feed crusher	511,000
Fork Lift	33,000
Contingency, 20% of Total Plant Cost	5,554,000
Total Plant Capital Cost Estimate	$33,871,000

OPERATING COST DETAILS

Reagents

Sulphuric acid consumption is based on preliminary testwork.

Sulphur dioxide consumption is based on preliminary testwork. A 10% loss was assumed.

Calcium oxide (lime) consumption was calculated to neutralize all acid added to the process (H_2SO_4+SO_2) less acid consumed by the lime in the ore (6.65% CaO).

Calcium chloride consumption was estimated at 5% of calcium chloride requirements (95% calcium chloride recovery).

Flocculant consumption was based on experience with other projects.

	Consumption kg/tonne ore	Annual Consumption tonne	Unit Cost $/tonne	Annual Cost $
H_2SO_4	180	30,960	130	4,025,000
SO_2	90	15,480	200	3,096,000
CaO	115	19,780	98	1,938,000
$CaCl_2$	10.7	1,840	398	732,000
Flocculant	0.25	43	3,100	133,000
Total Annual Reagent Cost				$9,924,000

Labour

Base labour rates have been assumed to be:

Hourly labour	$14.00/h
Shift foreman, supervision	$37,000/year
Payroll burden, 17% of salary or hourly rate	

Hydrometallurgical process plant labour requirements were estimated to be as follows:

Operating Labour Costs

Function, Mill Area	Personnel per shift	Total Personnel	Cost $/year
Stockpile loader	1	4	137,280
Crushing	1	2	68,640
Grinding-leaching	1	4	141,400
Acid filtration	1	4	137,280
Iron-gypsum precipitation tailings treatment	1	4	137,280
Manganese precipitation reagent recovery	2	8	274,560
Day labour	2	2	58,240
Sampler	1	4	116,480
Shift foreman	1	4	148,000
General foreman	1	1	40,000
Total		37	1,259,160
Salary burden, 17%			213,840
Total Annual Operating Labour Cost			1,473,000

Management and Administration

Function	Personnel per shift	Total Personnel	Cost $/year
Mill Superintendent		1	60,000
Mill Metallurgist		1	50,000
Plant Engineer		1	47,000
Powerhouse Engineer		1	40,000
Powerhouse Operator	1	4	140,000
Warehouse Foreman		1	37,000
Warehouse labourer	1	4	136,000
Laboratory Chemist		1	37,000
Laboratory Technician		2	68,000
Secretary		1	30,000
Total		17	645,000
Salary burden, 17%			110,000
Total Management and Administration Labour Cost			755,000
Total Annual Labour Cost			$2,228,000

Electrical Energy

Electrical power requirements are as follows:

	kW
Crushing and grinding	771.80
Neutral thickening and filtration	45.45
Leaching	150.75
Acid filtration	96.20
Iron-gypsum precipitation	105.25
Manganese precipitation; reagent recovery	69.95
Tailing neutralization	30.00
Reagent storage and distribution	24.70
Process air, vacuum and steam	189.00
Lighting allowance	120.00
Total	1,603.10

Cost of electrical energy assumed at \$0.05/kWh

Total annual energy 10.2 x 10^6 kWh
(assuming 80% usage)

Total Annual Electrical Energy Cost	\$508,000

Fuel Oil

Fuel oil is required for steam generation for the evaporator, for product drying and for the mobile equipment (front-end loader). The vapour discharge from the fourth evaporator stage will be used for leach and building heating.

Steam for evaporator (400 BTU/lb water)	34.1 x 10^6 BTU/h
Heat for dryer	5.5 x 10^6 BTU/h
Total heat	39.6 x 10^6 BTU/h
Fuel oil heat content	160,000 BTU/US gal
Fuel oil cost	\$0.258/L
Efficiency	80%

Fuel oil, annual process cost	\$2,396,000
Fuel oil, annual mobile equipment	\$ 30,000
Total Annual Fuel Oil Cost	\$2,426,000

Operating Supplies

Grinding Balls:	
Consumption	1 kg/tonne ore
Unit cost	$560/tonne
Total cost	$96,000/year
Liners:	
Consumption	0.075 kg/tonne ore
Unit cost	$1,650/tonne
Total cost	$21,000/year
Disc Bags:	
Consumption	580/year
Unit cost	$40 each
Total cost	$23,000/year
Filter Cloths:	
Consumption	100/year
Unit cost	$600 each
Total cost	$60,000/year
Total Annual Operating Supplies Cost	$200,000

Maintenance

Maintenance costs, including maintenance labour and supplies are 3.5% of the total plant capital cost estimate of $33,870,000.

Total Annual Maintenance Cost	$ 1,185,000

Summation	per year
Total Operating Cost	16,471,000
Contingency, 10% of Total Operating Cost	1,647,000
Total Plant Operating Cost Estimate	$18,118,000
Cost per kg Mn contained in produce	$1.81

APPENDIX II

EFFECTS OF PRE-CONCENTRATION ON THE STANDARD HYDROMETALLURGICAL PROCESS

PROCESS AND PLANT DESCRIPTION

This process alternative differs from the base case by having a bank of spiral concentrators located between the SAG and ball mills.

Run of mine ore is crushed and then ground in a larger SAG mill to a nominal 20 mesh grind which is suitable for gravity concentration. Primary cyclone overflow is passed through the spiral concentratora with the concentrate advancing to the secondary grinding circuit. Secondary grinding is completed in a smaller ball mill. The concentrator rejects are discharged to the tailing disposal area.

Gravity concentration is rather inefficient in that manganese recovery is only approximately 50% in approximately 30% of the original ore weight. However the concentration operation effectively rejects acid consumers in the ore resulting in a significant reduction in reagent requirement by the concentrate (50% reduction per kilogram of recovered manganese product). Concentrate manganese grade is approximately double the ore grade.

The process and plant circuit from secondary grinding through to product recovery do not differ from those of the base case with the exception of changes in size as required by the lower concentrate tonnage and solution flows.

Leaching takes place at elevated temperatures in agitated vessels with the addition of sulphuric acid and sulphur dioxide, and pregnant solution is recovered in the two stage acid filtration circuit.

The pregnant solution is treated with calcium chloride for gypsum removal and lime slurry for precipitation of other impurities.

The purified pregnant solution is treated with lime for manganese hydroxide precipitation. The product is recovered by thickening and filtraion and dried in a multi-hearth furnace. Calcium chloride is recovered in the reagent recovery circuit.

The sizes of the leaching, acid filtration, impurity precipitation and reagent recovery circuits are reduced because of lower tonnage throughput. The product circuit size remains unchanged.

The reagent recovery circuit, a multi-effect evaporation system used to concentrate the depleted calcium chloride solution is smaller as less solution must be evaporated to maintain the solution balance.

DESIGN CRITERIA

The basic design criteria for this process alternative is:

Ore Tonnage	-annual	325,207 tonne
	-average	41.01 tonne/h
	-design	45.16 tonne/h
Ore Grade		7.57% Mn
Pre-Concentration Design	-feed	45.16 tonne/h
	-concentrate	12.41 tonne/h
	-tailing	32.75 tonne/h
	-recovery manganese	52.28%
	-weight	27.48%
Production	-annual	10,000 tonne contained Mn
Product		$Mn(OH)_2$, 60%Mn
Recovery from Concentrate	-leaching	80.0%
	-overall	76.8%
Pregnant Solution Flow to Purification		18.6 tonne/h
Acid Consumption	-concentrate	180 kg/tonne
	-ore basis	49.4 kg/tonne ore
SO_2 Consumption	-concentrate	90 kg/tonne
	-ore basis	24.7 kg/tonne ore
Lime Consumption	-concentrate	181 kg/tonne
	-ore basis	49.6 kg/tonne ore
$CaCl_2$ Consumption	-concentrate	18.9 kg/tonne
	-ore basis	5.2 kg/tonne ore

The engineering design criteria for equipment sizing are as follows:

Crushing and Grinding

Crushing	-schedule	2 shifts per day 5 days per week
	-rate	94.84 tonne/h
Grinding	-schedule	24 h per day 7 days per week
SAG Mill	-rate	12.41 tonne/h
	-grind	nominal 20 mesh
	-energy	6.5 kWh/tonne
Classification		Cyclones

Spiral Concentration

Feed	-type	SAG mill discharge
	-rate	45.16 tonne/h
Concentrate	-annual rate	89,367 tonne
	-average rate	11.28 tonne/h
	-design rate	12.41 tonne/h
	-Mn grade	14.57%
	-Mn recovery	52.28%
	-weight recovery	27.48%
Tailing	-annual rate	235,840 tonne
	-average rate	29.78 tonne/h
	-design rate	32.75 tonne/h
	-Mn grade	5.04%

Neutral Thickening and Filtration

Design criteria as for Section 4.1, Appendix 1.

Leaching

Design criteria as for Section 4.1, Appendix I.

Acid Filtration

Pregnant Solution Flow to Purification 18.60 tonne/h

All other design criteria as for Section 4.1, Appendix IV.1

Iron-Gypsum Precipitation

Design criteria as for Section 4.1, Appendix I.

Manganese Precipitation and Reagent Recovery

Design criteria as for Section 4.1, Appendix I.

Tailing Neutralization

Design criteria as for Section 4.1, Appendix I.

Reagent Storage and Distribution

Design criteria as for Section 4.1, Appendix I.

Process Air, Vacuum and Steam

All design criteria as for Section 4.1, Appendix IV.1

MATERIAL BALANCE DETAILS

A detailed material balance, as prepared for the base case, was not necessary for this process alternative and was not calculated. Rather, only the process flow streams which would have a significant effect on equipment sizes and reagent consumption were determined.

The process flows estimated were:

	Solids tonne/h	Solution tonne/h
SAG mill feed	45.16	-
Spiral concentrate	12.41	-
Spiral tailings	32.75	-
Ball mill feed	12.41	-
Pregnant solution flow to purification	-	18.6
Iron-Gypsum precipitation solids	7.8	-
Solution to be evaporated for reagent recovery	-	20.4

CAPITAL COST DETAILS

Crushing and Grinding

The inclusion of concentration circuit has resulted in a larger primary grinding circuit and a smaller secondary circuit.

Equipment	No.	Size	Operating Power, kW	Unit Cost,$	Total Cost,$
Grizzly	1	610 mm	-	11,000	11,000
Coarse ore bin	1	3.6mx3.6mx6.8m	-	21,000	21,000
Hydraulic rock breaker	1	4.88m	22.6	64,000	64,000
Coarse ore feeder	1	990 mmx4.27m	18.6	40,000	40,000
Jaw crusher	1	610mmx200mm	111.9	124,000	124,000
Belt conveyor	3	760mmx30m	3x 34.8	53,333	160,000
Dust collector	1	11,500m /h	11.3	19,000	19,000
Belt scale	1	760 mm wide	0.2	8,000	8,000
Sump pump	1	-	3.8	8,000	8,000
SAG mill	1	-	500	1,110,000	1,110,000
Pump box	1	1.6mx1.6mx1.6m	-	10,000	10,000
Primary cyclone feed pumps	2	-	15	5,000	10,000
Primary cyclones	4	-	-	8,250	33,000
Ball mill	1	-	100	198,000	198,000
Pump box	1	1.6mx1.6mx1.6m	-	10,000	10,000
Secondary cyclone Feed pumps	2	-	5	3,500	7,000
Secondary cyclone	4	-	-	6,750	27,000
Total			892.8		$1,860,000

Total Area Cost - $1,860,000 x 3.50 = $6,510,000

Spiral Concentration

Equipment	No.	Size	Operating Power, kW	Unit Cost,$	Total Cost,$
Feed tank					
c/w agitator	1	2m x 2m	7.5	16,000	16,000
Spiral feed					
Pumps	2	-	15	5,000	10,000
Spirals	1	-	-	60,000	60,000
Spiral conc.					
Tank c/w agitator	1	2m x 2m	7.5	16,000	16,000
Conc. pumps	2	-	5	3,500	7,000
Spiral tailings					
Tank c/w agitator	1	2m x 2m	7.5	16,000	16,000
Tailings pumps	2	-	15	5,000	10,000
Total			57.5		135,000

Total Area Cost = $135,000 x 3.5 = $473,000

Neutral Thickening and Filtration

The cost was factored (6/10 Rule) from the base case cost, Section 4.1, Appendix I on the basis of feed tonnage to the circuits.

Total Area Cost = $\$1,362,000 \left(\frac{12.41}{23.90}\right)^{0.6}$ = $919,000

Leaching

The cost was factored (6/10 Rule) from the base case cost, Section 4.1, Appendix I on the basis of feed tonnage to the circuit.

Total Area Cost = $\$3,042,000 \left(\frac{12.41}{23.90}\right)^{0.6}$ = $2,053,000

Acid Filtration

The cost was factored (6/10 Rule) from the base case cost, Section 4.1, Appendix I on the basis of feed tonnage to the circuit.

Total Area Cost = $\$3,759,000 \left(\frac{12.41}{23.90}\right)^{0.6} =$ $\underline{\$2,537,000}$

Iron-Gypsum Precipitation

The costs of the solution handling portion of the circuit are a function of total solution volume (pregnant solution plus recirculating calcium chloride solution). The cost of the solids handling portion of the circuit is proportional to the solids tonnage (gypsum solids). Costs have been estimated on this basis using the 6/10 Rule and the base case costs, Secton 4.1, Appendix I.

Base case solution volume = 48.95 tonne/h

Present case solution volume = 30.60 tonne/h

Solution handling equipment costs - base case

Iron-gypsum precipitation tanks	$ 216,000
Discharge pumps	30,000
Iron-Gypsum thickener	339,000
Total Cost	$ 585,000

Solids handling equipment cost - base case

$914,000 - 585,000 = $ 329,000

Solution handling equipment cost - present case

Equipment cost = $\$585,000 \left(\frac{30.6}{48.95}\right)^{0.6} =$ 366,000

Base case solids flow = 9.73
Present case solids flow = 7.8

Solids handling equipment cost - present case
Equipment cost = $\$329{,}000\left(\frac{7.8}{9.73}\right)^{0.6} =$ \$288,000

Total area equipment cost = \$654,000

Total Area Cost = \$654,000 x 3.5 = \$2,289,000

Manganese Precipitation and Reagent Recovery

As for the iron-gypsum precipitation circuit, the appropriate equipment costs are a function of the solution volume and solids tonnage in the manganese precipitation and drying sections. The evaporator circuit costs are a function of solution volume evaporated. Costs have been factored from the base case costs, Section 4.1, Appendix I, on this basis using the 6/10 Rule.

Base case solution volume	45.56 tonne/h
Present case solution volume	27.2 tonne/h

Solution handling equipment - base case

Manganese precipitation tanks	\$ 108,000
Pump box	7,000
Thickener feed pumps	18,000
Manganese thickener	339,000
Total Cost	\$ 472,000

Solution handling equipment - present case
Equipment cost = $\$472{,}000\left(\frac{27.2}{45.56}\right)^{0.6} =$ \$ 346,000

Base case evaporated volume = 38.79 m^3/h
Present case evaporated volume = 20.44 m^3/h

Evaporator cost = $\$750{,}000 \left(\frac{20.4}{38.79}\right)^{0.6}$ = $ 511,000

Concentrate pumps	7,000
Total evaporator circuit	$ 518,000

The quantity of solids is the same in both cases and the solids handling circuit cost is:

Equipment cost = $2,091,000 - 472,000 - 757,000 = $ 862,000

Total area equipment cost	$1,726,000

Total Area Cost = $1,726,000 x 3.5 = $6,041,000

Tailing Neutralization

The costs are a further solid tailing and have been factored from the base case using the 6/10 Rule, Section 4.1, Appendix I.

Base case solids tonnage = 30.58 tonne/h
Present case solids tonnage = 19.1 tonne/h

$$\text{Total Area Cost} = \$452{,}000 \left(\frac{19.1}{30.58}\right)^{0.6} = \qquad \$\underline{\underline{\ \ 341{,}000}}$$

Reagent Storage and Distribution

Costs were assumed, the same as for the base case, Section 4.1, Appendix I.

Total Area Cost = $1,358,000

Process Air, Vacuum and Steam

The 3 - 500 BHP steam boilers were replaced by 3 - 300 BHP boilers.

The remainder of the equipment is the same.

Equipment cost = $464,000 - $263,000 + $195,000 = 396,000
Total Area Cost = $396,000 x 3.5 = $1,386,000

Summation

Total Plant Cost	$23,907,000
Frontend Loader to Feed Crusher	511,000
Fork Lift	37,000
Contingency, 20% of Total Plant Cost	4,781,000
Total Plant Capital Cost Estimate	$29,232,000

OPERATING COST ESTIMATE

Reagents

Sulphuric acid and sulphur dioxide consumptions were based on preliminary testwork. A 10% loss of sulphur dioxide was assumed.

Calcium oxide (lime) consumption was calculated to neutralize all acid added (H_2SO_4 and SO_2). There was no credit for acid consumed by ore, assuming lime in the ore was rejected by pre-concentration, as suggested by the preliminary testwork..

Calcium chloride consumption was estimated at 5% of requirements (95% recovery).

Flocculant consumption was based on requirements of other projects.

	Consumption kg/tonne Conc	Annual Consumption, tonne	Unit Cost $/tonne	Annual Cost $
H_2SO_4	180	16,090	130	2,092,000
SO_2	90	8,040	200	1,586,000
CaO	181	16,180	98	1,812,000
$CaCl_2$	18.9	1,690	398	673,000
Flocculant	0.25	23	3,100	71,000
Total Reagent Cost				$6,031,000

Labour

Labour requirements are as for the base case (Section 4.1, Appendix I) with the addition of one operator per shift for the additional circuits.

Additional labour cost = $137,280 x 1.17 (annual)= $ 161,000

Total Labour Cost (annual). =$2,228,000 + 161,000 = $2,389,000

Electrical Energy

Estimate electrical power requirements are as follows:

	kW
Crushing and grinding	892.8
Spiral concentration	57.5
Neutral thickening and filtration	31.0
Leaching	102.0
Acid filtration	65.0
Iron-gypsum precipitation	75.0
Manganese precipitation;reagent recovery	58.0
Tailing neutralization	23.0
Reagent storage and distribution	24.7
Process air, vacuum and steam	184.0
Lighting allowance	120.0
Total	1,633.0

Total annual energy = 10.4×10^6 kWh
(assuming 80% usage)

Total Electrical Energy Cost (annual)	$ 517,000

Fuel Oil

Steam for evaporator	18.0×10^6	BTU/h
Heat for dryer	5.5×10^6	BTU/h
Building heat	0.3×10^6	BTU/h
Total Heat	23.8 x 10	BTU/h

Fuel oil, process cost (annual)	$1,440,000
Fuel oil, mobile equipment (annual)	30,000
Total Fuel Oil Cost (annual)	$1,470,000

Operating Supplies

Grinding Balls:

Consumption	1 kg/tonne ore
Unit cost	$560/tonne
Tonnage	414,574 tonne/year
Total cost	$232,000/year

Liners:

Consumption	0.075 kg/tonne ore
Unit cost	$1,650/tonne
Tonnage	414,574 tonne/year
Total cost	$51,000/year

Disc Bags:

Consumption	580/year
Unit cost	$27 each
Total cost	$16,000/year

Filter Cloths:

Consumption	100/year
Unit cost	$400 each
Total cost	$40,000/year

Total Operating Supplies Cost (annual) $ 339,000

Mainentance

Maintenance costs (labour and materials) are 3.5% of the total plant capital cost estimate of $29,232,000.

Total Maintenance Cost (annual) = $1,023,000

Summation

	per year
Total Operating Cost	$11,769,000
Contingency, 10% of Total Operating Cost	1,177,000
Total Plant Operating Cost Estimate	$12,946,000
Cost per kg Mn contained in product	$1.30

APPENDIX III

EFFECTS OF HEAP LEACHING ON THE STANDARD HYDROMETALLURGICAL PROCESS

PROCESS AND PLANT DESCRIPTION

Heap leaching of the ore causes a radical change in the process up to the point of iron-gypsum precipitation. Manganese extraction by heap leaching results in the deletion of the grinding, acid leaching and acid filtration circuits of the base case.

Run-of-mine ore is crushed to minus 3/8 inch in a three stage crushing circuit consisting of a jaw crusher and two cone crushers. The final cone crusher is in a closed circuit with a vibrating screen which produces the final product.

Crushed ore is trucked to the heap leaching site and stacked by a front end loader. The ore is stacked on an impervious liner covering a prepared pad. Once the desired tonnage has been deposited on the pad, a piping and sprinkler network is installed to deliver leaching solution to the heap. The heap system is designed to hold two tiers of fresh ore, and the piping systems are recovered and re-installed for each tier.

Leaching solution, the leachant, with reagent added, is pumped through the piping network and sprinkled on the heap at a rate sufficient to saturate the heap. The leachant percolates down through the heap, drains into collection ditches and is stored in a holding pond. Heaps are leached in tandem (or countercurrently) to maintain a relatively steady grade in the pregnant solution. This arrangement is necessary because the extraction rate during the primary leach period is eight times faster than the extraction rate during the secondary period.

The partially depleted heap (Heap No. 2) is leached with circulating solution from the No. 2 Pond. Leaching reagents are added to the solution as it is circulated. A small portion of the solution from this pond is transferred to the No. 1 Pond system. Water and/or recycle process solution are added for make-up.

The newest heap in the tandem is leached with circulating solution from the No. 1 Pond. Again, leaching reagents are added to the solution as it is circulated. A small portion of the solution from this pond is advanced to the iron-gypsum precipitation circuit as pregnant aqueous solution.

As heaps (and/or tiers) become depleted, new heaps are brought on stream, and the countercurrent leaching scheme is maintained by making the oldest heap the second and the newest the first.

The pregnant aqueous solution is treated in the iron-gypsum precipitation circuit as described for the base case. Calcium chloride and lime are added to precipitate the impurities which are removed by thickening and filtration. The removed solids are treated in the tailing neutralization circuit.

Manganese hydroxide is precipitated with lime and recovered by thickening and filtration. The solids are dried in a multi-hearth furnace.

Spent calcium chloride solution is concentrated by evaporation to maintain a solution balance. The recovered reagent is recycled to process.

All process waste solution and solids (most iron-gypsum solids) are treated in the tailing neutralization circuit prior to discharge to the tailings basin. The depleted heaps are abandoned in place without further treatment.

In cases where vats might be used in place of heaps, the vats will hold a tonnage equal to a two tier heap pad. The sidewalls of the vats will be constructed from low grade or mine waste material. The vat will be lined with an impervious liner and a permanent piping network will be installed on top of the liner. Leach solution can be pumped into or withdrawn through this network. The vats will also be leached in tandem, and depleted vats will be abandoned.

DESIGN CRITERIA

The basic design criteria for this process alternative are:

Ore Tonnage	- annual	209,683 tonnes
	- average	26.48 tonne/h
	- design	29.12 tonne/h
Ore Grade		7.57% Mn
Production	- annual	10,000 tonne contained Mn
Product		$Mn(OH)_2$, 60% Mn
Leaching Period	- primary	20 days
	- secondary	20 days
	- total	40 days
Recovery	- leaching	65.0%
	- overall	63.0%
Leach Solution Bleed to Process		36.95 tonne/h
Acid Consumption		180 kg/tonne ore
SO_2 Consumption		90 kg/tonne ore
Lime Consumption		107 kg/tonne ore
$CaCl_2$ Consumption		10.6 kg/tonne ore

The engineering design criteria for the material balance and equipment sizing are as follows:

Crushing

Crushing	- schedule	2 shifts per day, 5 days per week
	- rate	82 tonne/h
	- stages	3
Product	- size	minus 3/8 inch
Screening		rod deck
Product Transfer		truck

Heap Leaching

Heap	- size, top tier	14,000 tonne
	- size, bottom tier	22,000 tonne
	- area	3720 m^2
Leaching	- method	tandem, 2 heaps
	- time, primary	20 days, continuous
	- time, secondary	20 days, continuous
	- time, total	40 days, continuous
Extraction Rate		
	- primary	2.90% per day
	- secondary	0.35% per day
	- average	1.63% per day
SO_2	- consumption	90 kg/tonne ore
	- loss	10%

Solution Handling

Leach Solution	- percolation rate	15-50 L/min m^2
	- flow per heap	360 m^3/h
Reagent Mixing		packed columns
Solution Ponds	- numbers	2
	- size	8700 m^3 (heap void volume + 25%)
Pregnant Solution Bleed to Purification		36.95 tonne/h

Iron-Gypsum Precipitation

Design criteria as for Section 4.1, Appendix 1.

Manganese Precipitation and Reagent Recovery

Design criteria as for Section 4.1, Appendix I.

Tailings Nuetraliztion

Design criteria as for Section 4.1, Appendix I.

Reagent Storage and Distribution

Design criteria as for Section 4.1, Appendix I.

Process Air, Vacuum and Stream

Design criteria as for Section 4.1, see Appendix I.

MATERIAL BALANCE DETAILS

A detailed material balance as prepared fro the base case, was not necessary for this process alternative and was not calculated. The process flow streams which have a significant effect on equipment sizes and operating cost were determined.

The process flows estimated were:

	Solids tonne/h	Solution tonne/h
Crushing plant feed	82	-
Design ore processing rate (24 h basis)	29.12	-
Heap leach solution flow	-	360
Pregnant solution flow to purification	-	36.95
Iron-gypsum precipitation solids	12.4	-
Solution to be evaporated for reagent recovery	-	40.9

CAPITAL COST DETAILS

Crushing

Equipment	No.	Size	Operating Power, kW	Unit Cost,$	Total Cost,$
Grizzly	1	610 mm	-	11,000	11,000
Coarse ore bin	1	3.6mx3.6mx6.8m	-	21,000	21,000
Hydraulic rock breaker	1	4.88m	22.6	64,000	64,000
Coarse ore feeder	1	990 mmx 4.27 m	18.6	40,000	40,000
Jaw crusher	1	610 mmx 203 mm	111.9	124,000	124,000
Belt conveyors	3	762 mmx30m	3x34.8	53,333	160,000
Belt scales	1	760 mm wide	0.2	8,000	8,000
Transfer conveyor belt	2	762mx15m	2x34.8	27,000	54,000
Dust collector c/w fan	1	11,530m^3/h	11.3	19,000	19,000
Sump pump	1	150 L/min	3.8	8,000	8,000
Vibrating grizzley	1	910 mmx1.5m	3.8	7,500	7,500
Standard cone crusher	1	910 mm	45	110,000	110,000
Vibrating screen	1	1020 mmx 3.6	5.6	21,500	21,500
Short head crusher	1	910 mm	56	113,000	113,000
Load-out bin	1	3.6 m x 3.6 m	-	16,000	16,000
Total			452.8		777,000

Total Area Cost = $777,000 x 3.5 $2,730,000

Heap Pads and Solution Ponds

Heap pads	2 installed cost	90,000
Solution ponds	2 installed cost	104,000
Heap pad piping	2 sets installed	76,000
Total		270,000
Electrical		27,000
Instrumentation		27,000
Field Expenses		35,000
Project management, engineering		97,000
Total Area Cost		$456,000

Solution Handling

Equipment	No.	Size	Operating Power, kW	Unit Cost,$	Total Cost,$
H_2SO_4/SO_2 tower	1	1.0 m x 4.5 m	-	12,000	12,000
H_2SO_4/SO_2 pumps	2		1x1.5	3,000	6,000
Reagent mix tank	2	3.5 m x 3.5 m	-	21,000	42,000
Pond recycle pumps	4	100 L/sec	2x45	16,000	64,000
Heap feed pumps	4	100 L/sec	2x38	12,000	48,000
Mill feed pumps	2	-	1x11	6,000	12,000
Total			148.5		184,000

Enclosures	46,000
Plant services	15,000
Site preparation	15,000
Field expenses	24,000
Project management, engineering	66,000
Total Area Cost	$527,000

Iron-Gypsum Precipitation

The cost of the solution handling portion of the circuit is the same as for the base case, Section 4.1, Appendix I, because solution flows are approximately the same.

The solid portion of the circuit has been fractured (6/10 Rule) using the base case costs. See Appendix II for equipment cost details.

Solution handling equipment cost	$ 585,000

Base case solids = 9.73 tonne/h
Present case solids = 12.4 tonne/h

Solids handling portion:

Equipment cost = $\$329,000 \left(\frac{12.4}{9.73}\right)^{0.6}$ =	$ 381,000
Total area equipment cost	$ 966,000
Total Area Cost = $966,000 x 3.5 =	$3,381,000

Manganese Precipitation and Reagent Recovery

The solid and solution flows are not significantly different from the base case. Costs for the manganese precipitation and drying cricuit cost will be the same.

Some additional water must be evaporated. Cost are factored, using the 6/10 Rule, and cost for Section 4.1, Appendix I..

Equipment cost less evaporator

=\$2,091,000 - \$750,000 = \$1,341,000

Base case evaporated volume = 38.79 m^3/h

Present case evaporated volume = 40.9 m^3/h

Evaporator cost = \$750,000 $\left(\frac{40.9}{38.79}\right)^{0.6}$ \$ 744,000

Total area equipment cost \$2,115,000

Total Area Cost = \$2,115,000 x 3.5 \$7,403,000

Tailing Neutralization

Costs are a function of solids flow and have been factored from the base case using 6/10 Rule and cost for Section 4.1, Appendix I).

Base case solids tonnage = 30.58 tonne/h

Present case solids tonnage = 12.4 tonne/h

Total Area Cost = \$452,000 $\left(\frac{12.4}{30.58}\right)^{0.6}$ = \$ 263,000

Reagent Storage and Distribution

Costs were as for the base case Section 4.1 Appendix I.

Total Area Cost \$1,358,000

Process Air, Vacuum and Steam

Equipment was adjusted to reflect deletion of the acid filters. Plant air will be used for process air because requirements are small.

Equipment	No.	Size	Operating Power, kW	Unit Cost, $	Total Cost, $
Vacuum pumps	2	1500 CFM	38	12,000	24,000
Steam boilers (complete)	3	500 BHP	20	-	263,000
Total			58		287,000

Total Area Cost = $287,000 x 3.5 $1,005,000

Summation

Total Plant Cost	$ 17,003,000
Front end loader - crusher feed	511,000
Truck - crushed product to heap	525,000
Front end loader - heap work	511,000
Fork lift	33,000
Contingency, 20% of Total Plant Cost	3,423,000
Total Plant Capital Cost Estimate	$22,116,000

OPERATING COST DETAILS

Reagents

Sulphuric acid and sulphur dioxide consumption were based on the preliminary testwork. A 10% loss of sulphur dioxide was assumed.

Calcium oxide (lime) consumption was calculated to neutralize all acid added (H_2SO_4 and SO_2) less the loss of SO_2 to atmosphere, and less the acid consumed by lime in the ore (6.65% CaO).

Calcium chloride consumption was estimated at 5% of requirements (95% recovery).

Flocculant consumption was based on requirement at other projects.

	Consumption kg/tonne ore	Annual Consumption tonne	Unit Cost $/tonne	Annual Cost $
H_2SO_4	180	37,740	130	4,906,000
SO_2	90	18,870	200	3,774,000
CaO	107	22,440	98	2,199,000
$CaCl_2$	10.6	2,230	398	886,000
Flocculant	0.05	10.5	3100	33,000
Total Reagent Cost				$10,998,000

Labour

Operating labour changes as compared with the base case (see Appendix I) as follows:

Delete 2 crusher loader operators	$ 63,640
Delete grinding-leaching operators	141,400
Delete acid filtration operators	137,280
Total	347,320
Salary burden, 17%	49,680
Total deletions (annual)	$ 397,000
Add 2 heap loader operators	68,640
Add 2 heap truck operators	68,640
Total	137,280
Salary burden, 17%	23,720
Total additions (annual)	$ 161,000

Total Annual Labour Cost = $2,228,000 - $397,000 + $161,000

$1,992,000

Electrical Energy	kw
Crushing	452.8
Solution handling	148.5
Iron-gypsum precipitation	112.0
Manganese precipitation; reagent recovery	69.95
Tailings neutralization	18.0
Reagent storage and distribution	24.7
Process air, vacuum and steam	58.0
Lighting allowance	120.0
Total	1,003.95

Total annual energy = 6.4×10^6 kWh/year (assuming 80% usage)

Total Electrical Energy Cost (annual) $318,000

Fuel Oil

Steam for evaporator	35.0 x 10^6 BTU/h
Heat for dryer	5.5 x 10^6 BTU/h
Total heat	41.5 x 10^6 BTU/h

Fuel oil, process cost	$2,511,000
Fuel oil - mobile equipment	90,000
Total Fuel Oil Cost	$2,601,000

Operating Supplies

Liners:	
Consumption	0.1 kg/tonne
Unit cost	$1,650/tonne
Total cost	$ 35,000

Filter Cloth:	
Consumption	35/year
Unit cost	$420 each
Total cost	$ 14,000

Total Operating Supplies Cost (annual)	$ 49,000

Heap Pad Preparation

Six pads are required per year.

Pad preparation cost, each	$ 45,000
Piping, moving and reinstall, each	17,000
Auxiliary, electrical, field services, engineering, each	56,000
Total per pad	$118,000

Total Heap Pad Preparation Costs = $118,000 x 6 = (annual)	$708,000

Maintenance

Maintenance costs (labour and material) are 3.5% of the total plant capita cost estimate of $22,116,000

Total Maintenance Cost (annual)	$ 774,000

Summation

Total Operating Cost (annual)	$17,440,000
Contingency, 10% of Total Operating Cost	$ 1,744,000
Total Plant Operating Cost Estimate (annual)	$19,184,000
Cost per kg Mn contained in product	$1.92

APPENDIX IV

EFFECT OF ELECTROWINNING ON THE STANDARD HYDROMETALLURGICAL PROCESS

PROCESS AND PLANT DESCRIPTION

Run-of-mine ore is crushed and then ground in a SAG ball mill circuit as described for the standard hydrometallurgical flowsheet. The subsequent neutral thickening and filtration circuit, leaching circuit, and filtration circuits are the same as and operate the same as those for the standard hydrometallurgical flowsheet (Section 4.1, Appendix I) with the exception that recycle anolyte is used for repulping leach feed, as well as repulping and washing in the filtration circuit. The recycle anolyte contains the acid generated during electrowinning, and its use in leaching (approximately 50% of the recycle) results in acid savings. The moisture content of the neutral filter cake limits the amount of anolyte that can be recycled.

The recovered pregnant solution advances to the impurity precipitation circuit. The acid filtration solids are discharged to tailing neutralization. Lime, ammonia and hydrogen sulphide are added to pregnant solution to precipitate iron, gypsum and base metals; to adjust solution pH; and to control the concentration of ammonium sulphate. Precipitated solids are removed by filtration and the solution is clarified in the clarification circuit ahead of electrowinning. The precipitated solids are discharged to tailings neutralization.

The purified and clarified pregnant solution is pumped into six banks of ten electrolytic cells. The cells are two compartment units where the anodes are separated from the cathodes by canvas diaphragms. The manganese metal is deposited in the stainless steel cathodes from the pH 8 solution. The solution flows through the diaphragm into the anode chamber where the solution is acidic. Ammonium sulphate concentrations are maintained at 130-140 g/L to stabilize cell performance.

The cathodes are removed from each cell about once every 48 hours, and the deposited metal is stripped from the starter sheets. The recovered metal is sold as recovered, without further processing.

The electrolytic recovery of manganese has a current efficiency of about 40%, resulting in the generation of a large quantity of heat. Since temperature control is critical, a very large volume of anolyte is recirculated to the anolyte treatment circuit for cooling in cooling towers. Manganese which is not deposited on the cathode is recycled for recovery via the leach circuit.

The circulating anolyte is passed through six cooling towers where the temperature drops from 40°C to 32°C. A large volume of the anolyte is evaporated, thus maintaining a solution balance in the overall circuit. The discharge from the cooling towers contains magnesium sulphate crystals which settle out in a thickener and which are removed with a small filtration circuit. The thickener overflow is recirculated to the tank house as the coolant. A bleed or recycle anolyte is diverted to leaching and acid filtration. The magnesium sulphate is pumped to the tailings neutralization circuit.

All process waste streams are sent to the tailing neutralization circuit for treatment with lime. The limed mixture is then fed to the ammonia recovery circuit. The ammonia must be removed because of stringent effluent standards. The ammonia is liberated by the high pH of the solution and stripped out by live steam injection. Removed ammonia is recycled to process and all tailings are discharged to the tailings impoundment area.

Plant auxiliaries such as the power house and reagent storage area are generally the same as for the base case with the addition of ammonia and hydrogen sulphide storage facilities.

DESIGN CRITERIA

The basic design criteria for this process alternative are:

Ore tonnage	- annual	172,000 tonne
	- average	21.70 tonne/h
	- design	23.90 tonne/h
Ore Grade		7.57% Mn
Production	- annual	10,000tonne Mn as manganese metal
Product		Manganese metal, 99% Mn
Recovery	- leaching	80.0%
	- overall	76.8%
Pregnant Solution Flow to Purification		36.95 tonne/h
Electrowinning Power Requirement		9.9 kWh/kg Mn
Acid Consumption		180 kg/tonne ore
SO_2 Consumption		90 kg/tonne ore
Lime Consumption		143 kg/tonne ore

The engineering design criteria for the material balance and equipment sizing are:

Crushing and Grinding

Design criteria as for Section 4.1, Appendix I.

Neutral Thickening and Filtration

Design criteria as for Section 4.1, Appendix I.

Leaching

Repulp Solution		recycle anolyte
	- percent of recycle	54.3
	- acidity	117 g/L
Leach Density	- minimum permitted	40% solids

Other design criteria as for Section 4.1, Appendix I.

Acid Filtration

Primary Repulp Solution		recycle anolyte
Secondary	- wash	water
	- wash ratio	as required

Other design criteria as for Section 4.1, Appendix I.

Impurity Precipitation

Base Metal		
Precipitation	- reagent	H_2S
	- metal dissolved in leach	0.36 kg/tonne ore as Cu
Solution pH	- final	8.0-8.5

Other design criteria as for the Iron-Gypsum Precipitation Circuit of Section 4.1, Appendix I.

Clarification

Pregnant Solution Flow		36.95 tonne/h
Filters	- type	pressure, downflow
	- number	4
	- media	anthracite
Clarified Solution	- solids content	10 ppm
Surge Tanks		before and after filters

Electrowinning

Feed solution	- flow	36.6 tonne/h
	- pH	8.0-8.5
	- Mn content	47 g/L
	- $(NH_4)_2.SO_4$ content	130 g/L
	- temperature	32-40°C
Energy	- requirement	9.9 kWh/kg Mn
	- efficiency	40%
Cooling Solution		recirculating anolyte
Cell	- size	1.22 m x 4.25 m
	- number of cathodes	20
	- type	two compartment canvas membrane
Cathode	- size	910 mm x 457 mm x 3 mm
	- type	stainless steel
	- spacing	200 mm
	- metal deposit thickness	250 mm
Anodes	- size	910 mm x 457 mm x 3 mm
	- type	Pb, Sn, Co alloy enclosed by the canvas membrane
Manganese Deposition Per Pass		80-90%

Anolyte Treatment

Anolyte Temperature	- in	40°C
	- out	32°C
Precipitated Solids		$MgSO_4$
Thickener Sizing		0.4 m/h/ upflow
Filter Sizing		assume same as gypsum filters
Cooling Method		cooling towers, forced air
Air	- temperature in	10°C
	- relative humidity in	70%
	out	90%

Tailing Neutralization and Ammonia Recovery

Tailing neutralization design criteria as for Section 4.1, Appendix I.

Ammonia Recovery	- process type	lime boil
	- number of stages	4
	- ammonia recovery	99%

Reagent Storage and Distribution

Sulphur Dioxide	- type	liquid H_2S
	- storage	1 tank
Ammonia	- type	liquid NH_3
	- storage	1 tank

Other design criteria as for Section 4.1, Appendix I.

Process Air, Vacuum and Steam

Design criteria as for Section 4.1, Appendix I.

MATERIAL BALANCE DETAILS

The solution balance determined for this process option are:

	Solids tonne/h	Solution tonne/h
Pregnant solution flow to impurity precipitation	-	36.95
Solution clarification	-	36.6
Solution feed to electrowinning	-	36.6
Anolyte recirculation	-	886
Anolyte recycle	-	24.2
Anolyte evaporated	-	10.6
Anolyte recycle to leach	-	11.5

CAPITAL COST DETAILS

Crushing and Grinding

Cost as for Section 4.1, Appendix I.

Total Area Cost $5,659,000

Neutral Thickening and Filtration

Cost as for Section 4.1, Appendix I.

Total Area Cost $1,362,000

Leaching

Cost as for Section 4.1, Appendix I.

Total Area Cost $3,042,000

Acid Filtration

Cost as for Section 4.1, Appendix I.

Total Area Cost $3,759,000

Impurity Precipitation

Costs estimated the same as for the iron-gypsum precipitation circuit, for Section 4.1, Appendix I.

Total Area Cost $3,199,000

Clarification

The cost was factored from circuit costs for a generally similar plant using the 6/10 Rule.

$$\text{Equipment cost} = \$387{,}000 \left(\frac{36.6}{84.1}\right)^{0.6} = \$\ 231{,}000$$

Total Area Cost = $231,000 x 3.5 $ 809,000

Electrowinning

Equipment	No.	Size	Operating Power, kW	Unit Cost,$	Total Cost, $
Feed surge tank	1	8 m dia. x 6 m		110,000	110,000
Feed discharge pumps	2	-	1 x 10	8,500	17,000
Electrolytic cells (6 lines 10 cells per line)	60	4.25 m x 1.22 m x 1.0 m		5,500	330,000
Catholyte feed tank	1	8 m dia. x 6 m		110,000	110,000
Catholyte pumps	12		6 x 5	4,500	54,000
Anolyte holding tank	1	8 m dia. x 6 m		110,000	110,000
Anolyte feed pumps	12	-	6 x 20	13,000	156,000
Anolyte return pumps	12	-	6 x 20	13,000	156,000
Anolyte discharge pumps	2		1 x 120	35,000	70,000
Cathode hoists & trollies	6	4 tonne		10,000	60,000
Cathode washing station	1	-		75,000	75,000
Cathode strip station	1	-		125,000	125,000
Cathode preparation station	1	-		75,000	75,000
Sludge handling system	1			75,000	75,000
Rectifiers	6	3200 hp		400,000	2,400,000
Total			400(1)		3,923,000

(1) excluding cells

Total Area Cost = $3,923,000 x 3.5 = $13,731,000

Anolyte Treatment

Equipment	No.	Size	Operating Power, kW	Unit Cost, $	Total Cost, $
Cooling towers c/w fans	6	600 ton	120	81,000	486,000
Recirculating pumps	6	-	120	13,000	78,000
Surge tank c/w agitator	1	4.5 m x 5.5 m	20	119,000	119,000
Thickener feed pumps	2	-	120	35,000	70,000
Thickener	1	60 m dia.	10	1,130,000	1,130,000
Thickener U/F pumps	2	-	7	15,000	30,000
Thickener O/F surge tank	1	8 m x 6 m	-	110,000	110,000
Anolyte return pumps	2	-	120	35,000	70,000
Filter	1	3.1 m x 3.1 m	2	72,000	72,000
Filter cake repulper c/w agitator	1	2 m x 2 m	7	16,000	16,000
Repulper pumps	2	-	7	7,500	15,000
Filtrate pumps	2	-	10	8,500	17,000
Vacuum receiver, moisture trap, seal tank	1 set		-	6,000	6,000
Total			543		2,219,000

Total Area Cost = $2,219,000 x 3.5 $ 7,767,000

Tailing Neutralization and Ammonia Recovery

Tailing neutralization circuit costs are as for Section 4.1, Appendix I.

Ammonia recovery circuit costs are as for a generally similar project.

Tailing neutralization cost	$ 427,000
Ammonia recovery cost	$ 452,000
Total Area Cost	$ 879,000

Reagent Storage and Distribution

Cost as for Section 4.1, Appendix I with the addition of sulphur dioxide and ammonia storage facilities.

Section 4.1 equipment cost	$ 388,000
H_2S storage tank, evaporator and pumps	89,500
NH_3 storage tank, evaporator and pumps	89,500
Total equipment cost	$ 567,000
Total Area Cost $567,000 x 3.5	$1,985,000

Process Air, Vacuum and Steam

Cost as for Section 4.1, Appendix I with replacement of the three 500 hp boilers with three 300 hp boilers.

Section 4.1 equipment cost	$ 464,000
Less 3 600 hp boilers	- 263,000
Plus 3 300 hp boilers	+ 195,000
Total equipment cost	$396,000
Total Area Cost = $396,000 x 3.5 =	$1,386,000

Summation

Total Plant Cost	$43,578,000
Front end loader to feed crusher	511,000
Fork lifts - 2	66,000
Contingency, 20% of Total Plant Cost	8,716,000
Total Plant Capital Cost Estimate	$52,871,000

OPERATING COST DETAILS

Reagents

Acid and sulphur dioxide consumptions are based on testwork. The quantity of fresh acid added to leach is reduced by the recycle of anolyte. The quantity of anolyte which can be used in leach is limited by the solution balance.

Lime consumption is based on neutralization of all added acid (and H_2S) less the lime present in the ore.

The hydrogen sulphide requirement is based on precipitation of the solubilized base metals (assume 100% dissolution during leach) during impurity precipitation.

Ammonia losses from the recovery circuit are based on projections for actual plant performance.

Flocculant consumption is based on requirements for generally similar projects.

	Consumption kg/tonne	Annual Consumption tonne	Unit Cost $/tonne	Total Cost (Annual) $
H_2SO_4	131	22,530	130	2,929,000
SO_2	90	15,480	200	3,096,000
CaO	147	25,280	98	2,477,000
H_2S	0.2	35	880	31,000
NH_3	0.09	16	410	6,000
Flocculant	0.25	43	3100	133,000
Total Reagent Cost				$8,672,000

Labour

Labour as for Section 4.1, Appendix I with deletions and additions.

Labour cost Section 4.1	$2,228,000
Delete - 2 manganese precipitation and recovery operators per shift 2 x 4 = 8 men x $34,320 x 1.17	- 321,000
Add - one tank house operator plus 1 helper per shift, 2 x 4 = 8 men	
- four cathode changers per shift 4 x 4 = 16 men	
- one cathode/anode preparation man per shift, 1 x 4 = 4 men	
- one cathode stripper per shift, 1 x 4 = 4 men	
- one anolyte treatment circuit operator per shift, 1 x 4 = 4 men	
Total operator additions = 36 men x $34,320 x 1.17	1,446,000
Add - one tankhouse foreman per shift 1 x 4 = 4 men x $37,000 x 1.17	173,000
Total Labour Cost (annual)	$3,526,000

Electrical Energy	kW
Crushing and grinding	771.80
Neutral thickening and filtration	45.45
Leaching	150.75
Acid filtration	96.20
Impurity precipitation	105.25
Clarification	50.00
Electrowinning (excluding cells)	400.00
Anolyte treatment	543.00
Tailing neutralization and ammonia recovery	57.00
Reagent storage and distribution	30.70
Process air, vacuum and steam	184.00
Lighting allowance	120.00
Total	2554.15

Energy per year (80% usage assumed)	16.2×10^6 kWh	
Electrolytic cells	99.0×10^6 kWh	
Total electrical energy	115.2×10^6 kWh/year	
Total Electrical Energy Cost (annual)		$5,760,000

Fuel Oil		
Leach heating	6.4×10^6 BTU/h	
Ammonia recovery	11.7×10^6 BTU/h	
Building heat	1.0×10^6 BTU/h	
Total heat	19.1×10^6 BTU/h	
Fuel oil, processed (annual)		$1,156,000
Fuel oil, mobile equipment cost		30,000
Total Fuel Oil Cost (annual)		$1,186,000

Operating Supplies

Grinding Balls:	
Cost as for Section 4.1, Appendix I	
Total cost	$96,000/year
Liners:	
Cost as for Section 4.1, Appendix I	
Total Cost	$21,000/year
Disc Bags:	
Cost as for Section 4.1, Appendix I	
Total Cost	$23,000/year
Filter cloth:	
Consumption, annual	110
Unit Cost	$600
Total Cost	$66,000/year
Cathodes:	
Consumption	2400 sheets/year
Unit Cost	$100
Total Cost	$240,000/year
Anodes:	
10% of cathode cost	
Total Cost	$ 24,000/year
Diaphragms:	
Consumption	9,600/year
Unit Cost	$20
Total Cost	$192,000/year
Total Operating Supplies Cost (annual)	$662,000

Maintenance

The maintenance cost (labour and materials) is 3.5% of the total plant capital cost estimate of $52,731,000.

Total Maintenance Cost	(annual)	$ 1,851,000

Summation	per year
Total Operating Cost	$21,657,000
Contingency, 10% of Total Operating Cost	2,166,000
Total Plant Operating Cost Estimate	$23,823,000
Cost per kg Mn metal produced	$2.63

APPENDIX V

EFFECT OF PRE-CONCENTRATION AND VAT LEACHING ON THE STANDARD HYDROMETALLURGICAL PROCESS

PROCESS AND PLANT DESCRIPTION

The process and plant description for this alternative is the same as for Sections 4.2 and 4.3 (Appendices II and III), with the exception that permanent leaching vats are used in place of the leaching heaps.

The leaching vats are constructed in a generally similar manner to leach heaps. Mine waste or low grade ore is used to form the sidewalls of the vats. The interior is lined with an impermeable membrane, and solution piping is placed on the bottom of the vat on top of the lining. This piping introduces and removes solution from the vats.

Run-of-mine ore is ground in the SAG mill and processed through the spiral concentration circuit. In this case the concentrate is pumped into the leaching vat through a distribution piping system. Mobile equipment for trucking concentrate is not necessary. The material in the vats is leached in the same manner as the heaps. Two vats are leached in tandem. The leach solution bleed is treated for manganese recovery as outlined in the previous descriptions.

DESIGN CRITERIA

The basic design criteria for this process alternative are:

Ore Tonnage	- annual	401,077 tonne/year
	- average	50.64 tonne/h
	- design	55.71 tonne/h
Ore Grade		7.57% Mn
Preconcentration:		
Design	- feed	55.71 tonne/h
	- concentrate	15.31 tonne/h
	- tailing	40.40 tonne/h
	- recovery - manganese	52.28%
	- weight	27.48%
Concentrate Tonnage	- annual	110,216 tonne/year
Production	- annual	10,000 tonne contained Mn
Product		$Mn(OH)_2$, 60% Mn
Recovery from Concentrate:		
	- leaching	65.0%
	- overall	63.0%
Leach Solution Bleed to Process		18.6 tonne/h
Acid Consumption	- concentrate	180 kg/tonne
	- ore basis	49.4 kg/tonne ore
SO_2 Consumption	- concentrate	90 kg/tonne
	- ore basis	24.7 kg/tonne ore
Lime Consumption	- concentrate	174 kg/tonne
	- ore basis	47.8 kg/tonne ore
$CaCl_2$ Consumption	- concentrate	17.2 kg/tonne
	- ore basis	4.7 kg/tonne ore

The engineering design criteria for equipment sizing and the material balance are as follows:

Crushing and Grinding

Crushing	- schedule	2 shifts per day 5 days per week
	- rate	120.0 tonne/h
Grinding	- type	SAG mill
	- schedule	24 hours per day 7 days per week
	- rate	55.71 tonne/h
	- grind	nominal 20 mesh
	- energy	9.0 kWh/tonne
Classification		cyclones

Spiral Concentration

Feed	- type	SAG mill discharge
	- rate	55.71 tonne/h
Concentrate	- annual rate	110,216 tonne
	- average rate	13.92 tonne/h
	- design rate	15.31 tonne/h
	- Mn grade	14.57%
	- Mn recovery	52.30%
	- weight recovery	27.48%
Tailings	- annual rate	290,861 tonne
	- average rate	36.72 tonne/h
	- design rate	40.40 tonne/h
	- Mn grade	5.04%

Vat Leaching

Vat	- size	40,000 tonne
	- area	3720 m^2
	- moisture content	10%
Leaching	- method	tandem, 2 vats
	- time, primary	20 days, continuous
	- time, secondary	20 days, continuous
	- time, total	40 days, continuous
Extraction Rate	- primary	2.90% per day
	- secondary	0.35% per day
	- average	1.63% per day
SO_2	- addition	90 kg/tonne conc.
	- loss	10%

Solution Handling

Leach solution	- percolation rate	15-50 L/min/m^2
	- flow per heap	360 m^3/h
Reagent mixing		packed column
Solution ponds	- number	2
	- size	8700 m^3
Pregnant solution bleed to precipitation		18.6 tonne/h

Iron-Gypsum Precipitation

Design criteria as for Section 4.1, Appendix I.

Manganese Precipitation and Reagent Recovery

Design criteria as for Section 4.1, Appendix I.

Tailing Neutralization

Design criteria as for Section 4.1, Appendix I.

Reagent Storage and Distribution

Design criteria as for Section 4.1, Appendix I.

Process Air, Vacuum and Steam

All design criteria as for Section 4.1, Appendix I.

MATERIAL BALANCE DETIALS

A detailed material balance, as prepared for the base case, was not necessary for this process alternative and was not calculated. The process flow streams which would have a significant effect on equipment size and reagent consumption were determined.

The process flows determined were:

	Solids tonne/h	Solution tonne/h
Crushing plant feed	120	-
SAG mill feed	55.71	-
Concentrate production	15.31	-
Vat leach solution flow	-	360
Pregnant solution flow to purification	-	18.6
Iron-gypsum precipitation solids	10.2	-
Solution to be evaporated for reagent recovery	-	20.4

CAPITAL COST DETAILS

Crushing and Grinding

The crushing plant size for the pre-concentration case, Section 4.2, Appendix II is suitable for this case.

The grinding section now includes a larger SAG mill and the deletion of the secondary grinding and classification circuits.

Area equipment cost	$1,860,000
Delete SAG mill	-1,110,000
Delete ball mill	- 198,000
Delete secondary classification	- 44,000
Net cost	$ 508,000
Add larger SAG mill	1,260,000
Total area equipment cost	$1,768,000

Total Area Cost = $1,768,000 x 3.5 $\underline{\$6,188,000}$

Spiral Concentration

The cost was factored (6/10 Rule) from the capital cost for the pre-concentration case, Section 4.2, Appendix II on the basis of ore tonnage.

$$\text{Total Area Cost} = \$473,000 \left(\frac{55.7}{45.2}\right)^{0.6} = \qquad \underline{\$\ 536,000}$$

Vat Leaching

The cost of the initial leaching vats and solution ponds was factored from the capital cost for the leaching pads Section 4.3, Appendix III on the basis of tonnage.

$$\text{Total Area Cost} = \$456,000 \left(\frac{40,000}{36,000}\right)^{0.6} = \qquad \underline{\$\ 486,000}$$

Solution Handling

The costs for the solution handling circuit for the heap leaching case, Section 4.3, Appendix III were assumed.

Total Area Cost	$ 527,000

Iron-Gypsum Precipitation

The quantity of solution handled is similar to the volume handled by the circuit for the pre-concentration case, Section 4.2, Appendix II and has been used.

Solution handling equipment cost	$ 366,000

The solids tonnage handled is similar to the tonnage for the base case, Section 4.1, Appendix I and has been used.

Solids handling equipment cost	$ 329,000	
Total area equipment cost	$ 695,000	
Total Area Cost = $695,000 x 3.5		$2,433,000

Manganese Precipitation and Reagent Recovery

Solution flow, solids tonnage and evaporated volume is similar to that for the pre-concentration case, Section 4.2, Appendix II.

Total Area Cost	$6,041,000

Tailing Neutralization

The solids tonnage treated in neutralization is 10.2 and costs have been factored (6/10 Rule) on the base case costs Section 4.1, Appendix I.

Total Area Cost = $\$452,000 \left(\frac{10.5}{30.58}\right)^{0.6} =$ $234,000

Reagent Storage and Distribution

Cost as for the base case Section 4.1, Appendix I.

Total Area Cost	$1,358,000

Process Air, Vacuum and Steam

Vacuum pump requirements are as for the heap leaching case, Section 4.3, Appendix III.

Steam boiler requirements are as for pre-concentration case, Section 4.2, Appendix II.

Vacuum equipment cost	24,000	
Steam boiler cost	195,000	
Total area equipment cost	$ 219,000	
Total Area Cost = $219,000 x 3.5		$ 767,000

Summation

Total Plant Cost	$18,570,000
Front end loader to feed crusher	511,000
Fork lift	33,000
Contingency, 20% of Total Plant Cost	$ 3,714,000
Total Plant Capital Cost Estimate	$22,828,000

OPERATING COST DETAILS

Reagents

Consumptions calculated on the same basis as for the pre-concentration and heap leaching cases (Sections 4.2 and 4.3, Appendices II and III).

	Consumption kg/tonne Conc	Annual Consumption tonne	Unit Cost $/tonne	Total Cost $
H_2SO_4	180	19,840	130	2,579,000
SO_2	90	9,920	200	1,984,000
CaO	174	19,180	98	1,880,000
$CaCl_2$	17.2	1,900	398	755,000
Flocculant	0.1	11.0	3100	34,000

Total Reagent Cost (annual) $7,232,000

Labour

Manpower requirements are as for the heap leaching case, Section 4.3, Appendix III with deletion of two heap loaders and 2 heap truck operators, but the addition of 4 operators for the process plant resulting in no net change.

Total Labour Cost (annual) $1,992,000

WD1

Electrical Energy	kW
Crushing and grinding	912.8
Spiral concentration	65.0
Solution handling	148.5
Iron-gypsum precipitation	81.0
Manganese precipitation; reagent recovery	58.0
Tailings neutralization	16.0
Reagent storage and distribution	24.7
Process air, vacuum and steam	44.0
Lighting allowance	120.0
Total	1,470.0

Total annual energy (assuming 80% usage)
9.3×10^6 kWh
Total Electrical Energy Cost (annual) $470,000

Fuel Oil

Requirements as for pre-concentration case, Section 4.2, Appendix II.

Total process heat	23.8×10^6 BTU/h	
Fuel oil, process	(annual)	$1,440,000
Fuel oil, mobile equipment	(annual)	30,000
Total Fuel Oil Cost	(annual)	$1,470,000

Operating Supplies

Grinding Balls:

Consumption	1 kg/tonne ore
Unit Cost	$560/tonne
Tonnage tonne/year	401,077
Total Cost	$225,000/year

Liners:

Consumption	0.1 kg/tonne ore
Unit Cost	$1,650
Tonnage	401,077
Total Cost	66,000/year

Filter cloth:

Consumption	35/year
Unit Cost	$400 each
Total Cost	14,000/year
Total Operating Supplies Cost (annual)	$305,000

Vat Preparation

Three vats are required per year.

Vat preparation cost	$ 48,000
Vat piping	41,000
Auxiliaries, electrical, field services, engineering	52,000
Total per vat	$141,000
Total Vat Preparation Cost- $141,000x3 (annual)	$423,000

Maintenance

Maintenance costs (labour and materials) are 3.5% of the total plant capital cost estimate of $22,828,000.

Total Maintenance Cost (annual)	$799,000

WD1

Summation

	Per Year
Total Operating Cost	$12,691,000
Contingency, 10% of Total Operating Cost	1,269,000
Total Plant Operating Cost Estimate	$13,960,000
Cost per kg Mn contained in product	$1.40

Summation

	per year
Total Operating Cost	$12,691,000
Contingency, 10% of Total Operating Cost	1,269,000
Total Plant Operating Cost Estimate	$13,960,000
Cost per kg Mn contained in product	$1.40

APPENDIX VI

PRE-CONCENTRATION AND VAT LEACHING FOR THE STANDARD HYDROMETALLURGICAL FLOWSHEET WITH ELECTROWINNING

PROCESS AND PLANT DESCRIPTION

The process and plant for this alternative from the jaw crusher through to vat leaching is the same as described for Section 4.5 in Appendix V.

Crushed ore is ground to a nominal 20 mesh grind (coarse grind) and concentrated in spiral concentrators. The concentrate is pumped into permanent vats and drained solution is removed by a piping network on the bottom of the vat. The vats drain to an estimated ten per cent moisture, such that all recycle anolyte can be used as leach solution. The recycle anolyte and recycling leach solution are saturated with SO_2 before entering the vats. Two vats are leached in tandem to maintain a steady manganese concentration in the high grade leach solution.

A bleed of high grade leach solution is treated with lime, ammonia and hydrogen sulphide as described for Section 4.4 in Appendix IV for impurity removal in the impurity precipitation circuit. The purified solution is clarified and fed into the electrolytic cells for metal recovery.

As before, anolyte is recirculated through the cells for waste heat removal and through the anolyte treatment circuit for cooling. A portion of the anolyte is evaporated during cooling and precipitated solid, mainly magnesium sulphate are removed from solution. The concentrated recycle anolyte is used in the leach circuit. The anolyte contains acid which is formed during metal recovery.

All solution and solid waste streams are treated in the tailings neutralization and ammonia recovery circuits for ammonia removal. The ammonia free tailings are discharged to the tailings basin. The recovered ammonia is recycled to process.

DESIGN CRITERIA

The basic design criteria for the process alternative are:

Ore Tonnage	-annual	401,077 tonne
	-average	50.64 tonne/h
	-design	55.71 tonne/h
Pre-concentration Design	-feed	55.71 tonne/h
	-concentrate	15.31 tonne/h
	-tailings	40.40 tonne/h
	-recovery, manganese	52.28%
	weight	27.48%
Concentrate Tonnage	-annual	110,216 tonne
Production	-annual	10,000 tonne contained Mn
Product		Mn metal, 99% Mn
Recovery from Concentrate		
	- leaching	65.0%
	- overall	63.0%
Leach Soltution Bleed to Process		36.95 tonne/h
Electrowinning Power Requirement		9.9 kwh/kg Mn
Acid Consumption	-concentrate	180 kg/tonne
	-ore basis	49.4 kg/tonne
SO_2 Consumption	-concentrate	90 kg/tonne
	-ore basis	24.7 kg/tonne ore
Lime Consumption	-concentrate	174 kg/tonne
	-ore basis	47.8 kg/tonne ore

The engineering design criteria for the process alternative are as follows:

Crushing and grinding

Design criteria as for Section 4.5, Appendix V.

Spiral Concentration

Design criteria as for Section 4.5, Appendix V

Vat Leaching

Recycle Anolyte Bleed to Leaching 100%

Other design criteria as for section 4.5, Appendix V.

Solution Handling

Pregnant Solution Bleed to Precipitation 36.95 tonne/h

Other design criteria as for Section 4.5, Appendix V.

Impurity Precipitation

Design criteria as form Section 4.4, Appendix IV.

Clarification

Design criteria as for Section 4.4, Appendix IV.

Electrowinning

Design criteria as for Section 4.4, Appendix IV.

Anolyte Treatment

Design criteria as for Section 4.4, Appendix IV.

Tailings Neutralization and Ammonia Recovery

Design criteria as for Section 4.4, Appendix IV.

Reagent Storage and Distribution

Design criteria as for Section 4.4, Appendix IV.

Process Air, Vacuum and Steam

Design criteria as for Section 4.1, Appendix I.

MATERIAL BALANCE DETAILS

The material balances determined for this process option are:

	Solids tonne/h	Solution tonne/h
Crushing plant feed	120	-
SAG mill feed	55.71	-
Concentrate production	15.31	
Anolyte recycle to leach	-	21.2
Vat leach solution flow	360	-
Pregnant solution flow to impurity precipitation	-	36.95
Impurity solids	10.2	
Solution feed to clarification	-	35.56
Solution feed to electrowinning	-	33.56
Anolyte recirculation	-	886
Anolyte recycle	-	21.2
Anolyte evaporated	-	10.6

CAPITAL COST DETAILS

Crushing and Grinding

Cost as for Section 4.5, Appendix V

Total Area Cost $6,188,000

Spiral Concentration

Cost as for Section 4.5, Appendix V

Total Area Cost $536,000

Vat Leaching

Cost as for Section 4.5, Appendix V

Total Area Cost $486,000

Solution Handling

Cost as for Section 4.3, Appendix III

Total Area Cost $527,000

Impurity Precipitation

Cost as for Section 4.4, Appendix IV

Total Area Cost $3,199,000

Clarification

Cost as for Section 4.4, Appendix IV

Total Area Cost $809,000

Electrowinning

Cost as for Section 4.4, Appendix IV

Total Area Cost $13,731,000

Anolyte Treatment

Cost for Section 4.4, Appendix IV

Total Area Cost	$ 7,767,000

Tailing Neutralization and Ammonia Recovery

Cost factored on the basis of solids treated using 6/10 Rule and Section 4.4, Appendix IV costs.

Total Area Cost = $\$879{,}000 \left(\frac{10.2}{30.0}\right)^{0.6} =$ $ 455,000

Reagent Storage and Distribution

Cost as for Section 4.4, Appendix IV

Total Area Cost	$ 1,985,000

Process Air, Vacuum and Steam

Air and vacuum requirement as for Section 4.3, Appendix III, plus three 300 HP boilers

Vacuum pumps, 2 @ 1500 CFM	$24,000	
Boilers, 3 @ 300 HP	130,000	
Total equipment cost	$154,000	
Total Area Cost = $154,000 x 3.5 =		$ 539,000

Summation

Total Plant Cost	$36,222,000
Front end loader to feed crusher	511,000
Forklifts, 2	66,000
Contingency, 20% of Total Plant Cost	7,244,000
Total Plant Capital Cost Estimate	$44,043,000

OPERATING COST DETAILS

Reagents

Acid and sulphur dioxide consumptions are based on testwork. The low moisture content (10%) of coarsely ground concentrate, and the solution balance permit complete recycle of anolyte to vat leaching.

Lime consumption is based on neutraliation of all acid (and H_2S) added less the SO_2 loss.

The hydrogen sulphide and ammonia consumption calculated as for Section 4.4, Appendix IV.

Flocculant consumption is based on consumption for generally similar projects.

	Consumption kg/tonne Conc.	Annual Consumption tonne	Unit Cost $/tonne	Total Cost $
H_2SO_4	18	1,980	130	257,000
SO_2	90	9,920	200	1,984,000
CaO	174	19,180	98	1,880,000
H_2S	0.31	34	880	30,000
NH_3	0.11	12	410	5,000
Flocculant	0.1	11	3,100	33,000
Total Reagent Cost			(per year)	$4,189,000

Labour

Cost as for Section 4.5, Appendix V with the additions listed below.

Cost for Section 4.5	$1,919,000
add- one operator per shift for the tailing neutralization - ammonia recovery circuit, 1x4 = 4 men @ $34,320 x 1.17	161,000
add- 40 men required for the electrowinning circuit as for Section 4.4, Appendix IV 40 men total cost of	$1,619,000
Total Labour Cost (annual)	$3,699,000

Electrical Energy	kW
Crushing and grinding	912.80
Spiral concentrate	65.00
Solution handling	148.50
Impurity precipitation	105.25
Clarification	50.00
Electrowinning (excluding cells)	400.00
Anolyte treatment	543.00
Tailing neutralization & Ammonia Recovery	30.00
Reagent storage & distribution	30.70
Process air vacuum and steam	53.00
Lighting allowance	120.00
Total	2458.25

Energy per year (80% usage assumed) = 15.6 x 10^6 kWh

Electrolytic cells 99.0 x 10^6 kWh

Total energy 114.6x10^6 kWh/year

Total Electrical Energy Cost (annual) $5,730,000

Fuel Oil

Process steam	$6.4x10^6$ BTU/h	
Building heating	$1.0x10^6$ BTU/h	
Ammonia recovery	$3.1x10^6$ BTU/h	
Total heat	10.5×10^6 BTU/h	
Fuel Oil Cost (annual)		$ 635,000
Fuel Oil mobile equipment costs (annual)		30,000
Total Fuel Oil Cost (annual)		$ 665,000

Operating Supplies

Grinding Balls:

Cost as for Section 4.5, Appendix V

Total cost	(annual)	$ 225,000

Liners

Cost as for Section 4.5, Appendix V

Total cost	(annual)	$ 66,000

Filter Cloth

Consumption		45/year
Unit cost		$400.00 each
Total cost	(annual)	$ 18,000

Cathodes

Cost as for Section 4.4, Appendix IV

Total cost	(annual)	$ 240,000

Anodes

Cost as for Section 4.4, Appendix IV

Total cost	(annual)	$ 24,000

Diaphragms

Cost as for Section 4.4, Appendix V

Total cost	(annual)	$ 192,000
Total Operating Supplies Cost	(annual)	$ 765,000

Vat Preparation

Cost as for Section 4.5, Appendix V

Total Vat Preparation Cost	(annual)	$ 423,000

Maintenance

Maintenance cost (labour and materials) are 3.5% of the total plant capital cost estimate of $44,043,000

Total Maintenance Cost	(annual)	$1,542,000

Summation

	Per Year
Total Operating Cost	$17,013,000
Contingency, 10% of Total Operating Cost	1,701,000
Total Plant Operating Cost Estimate	$18,714,000
Cost per kg Mn metal produced	$1.87

APPENDIX VII

PRECONCENTRATION AND VAT LEACHING FOR THE STANDARD HYDROMETALLURGICAL FLOWSHEET WITH SOLVENT EXTRACTION FOR PRODUCTION OF A HIGH PURITY CHEMICAL PRODUCT

PROCESS AND PLANT DESCRIPTION

The pre-concentration and vat leaching processes for the solvent extraction alternative will be the same as described in Appendix V.

The leach solution bleed from the vats is clarified in downflow pressure sand filters prior to solvent extraction. At this time, the di-(2-ethyehexyl) phosphoric acid (D2HPA) extractant is being considered as a possible extractant for the proposed solvent extraction process.

The extractant is dissolved in kerosene which also contains a suitable modifier to inhibit emulsion formation. Extraction of manganese from the pregnant aqueous will take place in mixer-settlers. Prior to stripping the manganese the organic may be treated or washed for removal of some impurities. The loaded organic will be advanced to stripping, where a suitable reagent will strip the manganese from the solvent. At this time it is assumed that an acid will be used for stripping. The net result of acid stripping is the release of hydrogen ions (acidification) into the aqueous raffinate solution. This provides acid for vat leaching. The barren organic which might be scrubbed for impurity removal is returned to the extraction portion of the circuit. Standard solvent extraction equipment will be used for extraction, washing, stripping and scrubbing.The aqueous raffinate solution is directed to heap leaching and to the tailings neutralization circuit for treatment. At this time, the tailings neutralization circuit provided for the standard flowsheet is considered adequate for this process alternative.

The purified strip solution containing the manganese is directed to the product purification circuit for product recovery.

The manganese is precipitated using a suitable reagent, filtered, dried and stored for shipment. The process and circuit details will reflect the selection of the stripping and precipitation reagents.

The solution bleed(s) from the precipitation circuit will be discharged into the reagent recovery facilities. The recovery processes and facilities will be designed for the specific reagents being recovered. Recovered reagents will be recycled to the solvent extraction stripping circuit and the product precipitation circuit.

Conventional equipment will be used for the precipitation and product drying circuits, and the reagent recovery facilites. Equipment will be modified as necessary to suit the specific conditions.

At this time stoichiometric quantities (Mn basis) of sulphuric acid and lime have been allowed for to permit reagent recovery. In addition sufficient fuel oil has been allowed for to permit evaporation of the same volume of solution as for the calcium chloride route in Section 4.5, Appendix V. These allowances are more than adequate to reflect possible reagent and reagent recovery costs.

DESIGN CRITERIA

The basic design criteria for this program alternative are:

Ore Tonnage	- annual	401,077 tonne/year
	- average	50.64 tonne/h
	- design	55.71 tonne/h
Ore Grade		7.57% Mn
Preconcentration Design	- feed	55.71 tonne/h
	- concentrate	15.31 tonne/h
	- tailing	40.40 tonne/h
	- recovery, manganese	52.28%
	weight	27.48%
Production	- annual	10,000 tonne contained Mn
Product		Chemical precipitate 1% listed impurities
Recovery from Concentrate	- leaching	65.0%
	- overall	63.0%
Leach Solution Bleed to Process		18.6 tonne/h
Acid Consumption	- concentrate	180 kg/tonne
	- ore basis	49.4 kg/tonne ore
SO_2 Consumption	- concentrate	90 kg/tonne
	- ore basis	24.7 kg/tonne ore
Lime Consumption	- concentrate	128 kg/tonne
	- ore basis	35.2 kg/tonne ore

The engineering criteria for equipment sizing and the material and energy balance are as follows:

Crushing and Grinding

Design criteria as for Section 4.5, see Appendix V.

Sprial Concentrator

Design criteria as for Section 4.5, see Appendix V.

Vat Leaching

Design criteria as for Section 4.5, see Appendix V.

Solution Handling

Design criteria as for section 4.5, see Appendix V.

Clarification

Pregnant Solution Flow	18.6 tonne/h
Number of Filters	3
Clarified Solution- solids content	<10ppm
Surge Tanks	before and after filters

Solvent Extraction

Pregnant Solution Flow	18.6 tonne/h
Extractant (example)	D2HPA
Carrier	Kerosene
Modifier (example)	TBP
Extraction Stages	6
Stripping Stages	4
After Settlers	2
Impurity Scrub Stage	4

Manganese Precipitation and Reagent Recovery

Design criteria as for Section 4.1, see Appendix I.

Tailings Neutralization

Design criteria as for Section 4.1, see Appendix 1.

Reagent Storage & Distribtuion

Design criteria as for Section 4.1, see Appendix 1.

Process Air Vacuum and Steam

Design criteria as for Section 4.1, see Appendix I.

MATERIAL BALANCE DETAILS

A detailed mass balance was not calculated for this alternative. The mass balance which was estimated was based on the balances for Sections 4.5 (Apendix V). The process steams which were estimated were:

	Solids tonne/h	Solution tonne/h
Crushing plant feed	120	-
SAG mill feed	55.71	-
Concentrate production	15.31	-
Vat leach solution flow	-	360
Pregnant solution flow to purification	-	18.6
Tailing neutralization solids	8.04	-
Raffinate bleed to tailing neutralization		9.3

CAPITAL COST DETAILS

Crushing and grinding

Cost as for Section 4.5, Appendix V.

Total Area Cost $6,188,000

Spiral Concentration

Cost as per Section 4.5, Appendix V.

Total Area Cost $ 536,000

Vat Leaching

Cost as for Section 4.5, Appendix V

Total Area Cost $ 486,000

Solution Handling

Cost as for Section 4.3, Appendix III

Total Area Cost $ 527,000

Clarification

The capital cost for the clarification circuit was factored (6/10 Rule) on the basis of pregnant aqueous flow from the capital cost for a clarification circuit for a generally similar solvent extraction process.

$$\text{Total Area Cost} = \$1,355,000 \left(\frac{18.6}{84.1}\right)^{0.6} = \$\ 550,000$$

Solvent Extraction

The capital cost for the solvent extraction circuit was factored (6/10 Rule) on the basis of pregnant aqueous flow from the capital cost for a generally similar process. The circuit used for the calculation was relatively complex in comparison with standard solvent extraction circuits. The estimated cost was also adjusted for production volume.

$$\text{Total Area Cost} = \$6,963,000\left(\frac{18.6}{84.1}\right)^{0.6} = \qquad \$2,820,000$$

Manganese Precipitation and Reagent Recovery

Cost as for Section 4.2, see Appendix II.

Total Area Cost $6,041,000

Tailings Neutralization

The cost was factored on the solids tonnage treated in the circuit using the 6/10 Rule and the cost for the base case, Section 4.1, Appendix I.

$$\text{Total Area Cost} = \$452,000\left(\frac{8.04}{30.58}\right)^{0.6} = \qquad \$203,000$$

Reagent Storage and Distribution

Cost as for Section 4.1, Appendix I

Total Area Cost $1,358,000

Process Air, Vacuum and Steam

Cost as for Section 4.5, Appendix V.

Total Area Cost $767,000

Summation

Total Plant Cost	$19,476,000
Front end loader to feed crusher	511,000
Forklift	33,000
Contingency, 20% of Total Plant Cost	3,895,000
Total Capital Cost Estimate	$23,915,000

OPERATING COST DETAILS

Reagents

The cost of the SX reagents (extractant, kerosene and modifier) were based on costs for a generally similar process. Costs were calculated proportional to pregnant aqueous flow.

The cost of the stripping-precipitation reagent were calculated on the same basis with allowances for reagent recovery. Allowance for sulphuric acid addition and lime neutralization are included.

	Consumption Concentrate kg/tonne	Annual Consumption tonne	Unit Cost $/tonne	Annual Cost $
H_2SO_4	99	10,910	130	1,418,000
SO_2	90	9,920	200	1,984,000
CaO	128	14,110	98	1,383,000
Flocculant	0.05	5	3100	17,000
SX Reagents	-	-	-	100,000
Stripping, Precip. Reagents	-	-	-	4,600,000
Total Reagent Cost			(annual)	$9,502,000

Labour

Cost as for Section 4.5 plus one operator per shift, for 4 shifts in the solvent circuit.

Labour cost as per Section 4.5, Appendix 4.5	(annual)	1,992,000
4 operators plus overheads		161,000
Total Labour Cost	(annual)	$2,153,000

Vat Preparation

Cost as for Section 4.5, See Appendix 4.5 for three vats per year

Total Vat Preparation Cost	(annual)	$ 423,000

Electrical Power

Cost as for Section 4.5, Appendix 4.5 plus 800,000 kWh/year for the solvent extraction circuit.

Electrical energy = $9.3 \times 10^6 + 0.8 \times 10^6$ = 10.1×10^6 kWh/year

Total Electrical Energy Cost (annual) $ 510,000

Fuel Oil

Fuel oil is used for building heat and assumed for reagent recovery.

Heat requirements as for Section 4.5, Appendix 4.5

Heat Requirement= 23.8×10^6 BTU/h

Total Fuel Oil Cost (annual) $1,470,000

Operating Supplies

Cost as for Section 4.5, Appendix 4.5 but delete the iron-gypsum filter cloth. $305,000 - $7,000

Total Operating Supplies Cost (annual) $ 298,000

Maintenance

35% of total plant capital cost estimate at $23,915,000

Total Maintenance Cost (annual) $ 837,000

Summation

	Per Year
Total Operating Cost	$15,193,000
Contingency, 10% of Total Operating Cost	1,519,000
Total Plant Operating Cost Estimate	$16,712,000
Cost per kg Mn, Contained in Product	$1.67

APPENDIX VIII

CONVENTIONAL HYDROMETALLURGICAL PROCESS FOR $MnSO_4$

PROCESS PLANT DESCRIPTION

This process alternative produces an agricultural grade manganese sulphate product ($MnSO_4$).

The process flowsheet from crushing through to impurity precipitation is identical to the flowsheet for the base case as described in Appendix I, with the exception that mother liquor is used for grinding dilution, acid filter interstage repulping and filter wash. The mother liquor originates in the manganese crystallization circuit and contains approximately 8 g/L manganese sulphate. Recycle throughout the circuit is necessary to maintain overall recovery.

An impurity precipitation circuit rather than a straight iron hydrolysis circuit was selected because lime pH adjustment and precipitation of solids removes most soluble metals. In addition the near neutral pH of the resultant purified pregnant solution is preferable for further treatment because less expensive (less corrosion resistance) metals can be used for the high temperature service in the crystallization circuit, and the neutral solution can be recycled to grinding thus maintaining recovery levels.

The purified pregnant solution is heated in two heat exchangers prior to entering the autoclave crystallization. In the first exchanger the solution is heated by the flashed vapour from the autoclave letdown tank. In the second exchanger the temperature is raised to 140°C with steam. At this temperature the solubility limit for the concentration level of manganese sulphate in solution is slightly higher than the actual solution concentration. Spare exchangers are provided because of expected scale formation. All wetted metal parts are 304 stainless steel which should be suitable for the near neutral solution.

Heating to the final crystallization temperature of 170°C is done in jacketed, steam heated autoclaves. The manganese sulphate crystalizes as $MnSO_4.H_2O$. Four autoclaves are provided with one being a spare autoclave and also a spare flash tank when required.

The autoclaves discharge into the flash tank for slurry cooling by evaporation. The vapours are used for solution preheating. The $MnSO_4.H_2O$ slurry is processed through a standard thickener-filter circuit for crystal recovery.

The recovered crystals are dried in a directly fixed multi-heated dryer at 300°C. Water of hydration is released and the final product, $MnSO_4$ is stored in a bin for bulk shipment.

DESIGN CRITERIA

The basic design criteria for this conventional hydrometallurgical process are:

Ore Tonnage	- annual	172,000 tonne
	- average	21.7 tonne/h
	- design	23.9 tonne/h
Ore Grade		7.57% Mn
Production	- annual	10,000 tonne contained Mn
Product	- $MnSO_4$	
Recovery	- leaching	80.0%
	- overall	76.8%
Pregnant Solution Flow to Purifiaction		36.95 tonne/h
Acid Consumption		180 kg/tonne ore
SO_2 Consumption		90 kg/tonne ore
Lime Consumption		56 kg/tonne ore

The engineering design criteria for equipment sizing and cost estimating are as follows:

Crushing and Grinding

Design criteria as for Section 4.1, Appendix I.

Neutral Thickening and Filtration

Design criteria as for Section 4.1, Appendix I.

Leaching

Design criteria as for Section 4.1, Appendix I.

Acid Filtration

Design criteria as for Section 4.1, Appendix I.

Iron-Gypsum Precipitaion

Design criteria as for Section 4.1, Appendix I.

Manganese Sulphate Crystallization and Product Drying

Primary heating	- solution	purified pregnant solution
	- equipment	tube and shell heat exchangers
	- heating fluid	condensing vapours from flash tank
	- heat recovery	80% overall
	- heat transfer co-efficent	400 BTU/(^{o}F).(ft^{2}).(h)
Secondary Heating	- solution	purified pregnant solution
	- heating fluid	500 psi steam
	- heat transfer co-efficient	400 BTU (^{o}F).(ft^{2}).(h)

Crystallization	- equipment	jacketed autoclaves
	- temperature	170°C
	- heating fluid	500 psi steam
	- terminal solubility	± 8 g/L $MnSO_4$
	- solution pH	± 7
	- heat transfer co-efficient	150 BTU(°F).(ft^2).(h)
Cooling	- equipment	let down tank
	- method	vapourization of solution
Product	- solid	$MnSO_4 \cdot H_2O$
	- recovery	thickening and filtration
Thickener	- sizing	0.4 m/h upflow
	- underflow density	35% solids
Filter	- sizing	270 kg/h/m^2
	- discharge moisture	50%
	- filtrate (mother liquor)	recycle to grinding and acid filtration
Drying	- product	$MnSO_4$
	- temperature	300°C
	- moisture content	1%
	- equipment	Direct fired multi-hearth

Tailing Neutralization

Design criteria as for Section 4.1, Appendix I.

Reagent Storage & Distribution

Design criteria as for Section 4.1, Appendix I.

Process Air, Vacuum and Steam

Design criteria as for Section 4.1, Appendix I.

MATERIAL BALANCE DETAILS

Material balance details suitable for equipment sizins and reagent and energy requirement estimates are as follows:

	Solids tonne/h	Solution tonne/h
SAG mill feed	23.90	-
Pregnant solution to purification	-	36.95
Impurity solids precipitated	6.6	-
Pregnant solution to crystallization	-	36.5
Vapour loss	-	4.6
Mother liquor recycle to grinding	-	3.3
Mother liquor recycle to acid filtration	-	18.0

CAPITAL COST DETAILS

Crushing and Grinding

Cost as for Section 4.1, Appendix I.

Total Area Cost $ 5,659,000

Neutral Thickening and Filtration

Cost as for Section 4.1, Appendix I.

Total Area Cost $ 1,362,000

Leaching

Cost as for Section 4.1, Appendix I.

Total Area Cost $ 3,042,000

Acid Filtration

Cost as for Section 4.1, Appendix I

Total Area Cos $ 3,759,000

Impurity Precipitation

Cost of the solid handling portion of the circuit was determined by factioning (6/10 Rule) cost as for Section 4.1, Appendix III

Solid handling porton:

Equipment cost $=\$329{,}000\ \left(\frac{6.6}{9.73}\right)^{0.6} =$ $ 261,000

Solution handling portion

Equipment cost $= \$505{,}000\ \frac{(36.6)}{(45.56)}^{0.6} =$ $ 447,000

Total equipment cost $ 703,000

Total Area Cost = $703,000 x 3.5 $ 2,461,000

Manganese Sulphate Crystallization and Product Drying

Product Drying

Equipment list

Equipment	No.	Size	Operating Power kW	Unit Cost $	Total Cost $
Primary heat exchanges	2	600 ft^2	-	37,000	74,000
Secondary heat exchanges	2	50 ft^2	-	5,000	10,000
Condensate tanks	2	0.8mx1.5m	-	2,000	4,000
Condensate pumps	4	-	2x2	1,500	6,000
Autoclave feed pump	2	-	35	23,000	46,000
Autoclave crystallizers	4	2.4mx2.4m	3x5	203,000	812,000
Flash tank	1	2.4mx2.4m	-	81,000	81,000
Discharge pumps	2	-	7	7,000	14,000
Thickener	1	22m	4	339,000	339,000
Thickener U/F pumps	2	-	11	15,000	15,000
Drum filter	1	$10m^2$	1	70,000	70,000
Vacuum system (set)	1	-	-	6,000	6,000
Filtrate pumps	2	-	6	3,500	7,000
Dryer	1	-	10	648,000	648,000
Conveyor belt	1	914mx15m	8	50,000	50,000
Dust collector	1	-	3	15,000	15,000
Product silo	1	5mx7m	-	37,000	37,000
Pump box	1	2mx2m	-	7,000	7,000
Solution recycle pump	2		10	5,000	5,000
Total		111			$2,246,000

Total Area Cost = $2,246,000 x 3.5 = $7,861,000

Tailing Neutralization

Cost as for Section 4.1, jAppendix I.

Total Area Cost	$ 452,000

Reagent Storage and Distribution

Cost as for Section 4.1, Appendix I.

Total Area Cost	$ 1,358,000

Process Air, Vacuum and Steam

Cost as for Section 4.2, Appendix II

Total Area Cost	$ 1,386,000

Summation

Total Plant Cost	$27,340,000
Front end loader to feed crusher	$ 511,000
Forklift	$ 33,000
Contingency, 20% of Total Plant Cost	$ 5,568,000
Total Plant Capital Cost Estimate	$33,352,000

OPERATING COST DETAILS

Reagents

Acid, sulphur dioxide and flocculant requirements calculated on same basis as Section 4.1, Appendix 1.

Lime requirements based on neutralization of all acid and SO_2 added less acid neturalization by lime in the ore and acid kused to $MnSO_4$.

Consumption kg/tonne ore		Annual Consumption, tonne	Unit Cost $/tonne	Annual Cost $
H_2SO_4	180	30,960	130	4,025,000
SO_2	90	15,480	200	3,096,000
CaO	56	9,630	98	944,000
Flocculant	0.25	43	3,100	133,000
Total Reagent Cost				$8,198,000

Labour

Costs as for Section 4.1, Appendix 1.

Total Labour Cost (annual)	$2,228,000

Electrical Energy	kW
Crushing and grinding	771.80
Neutral thickening and filtration	45.45
Leaching	150.75
Acid filtration	96.20
Impurity precipitation	88.00

Manganese sulphite crystallization and product drying	111.00
Tailing neutralization	30.00
Reagent storage and distribution	24.70
Process air, vacuum and steam	189.00
Lighting allowance	120.00
Total	1,626.90

Total annual energy = 10.6 $x^{10.6}$ kwh/year
Assuming 80% usage)

Total Electrical Energy Cost (annual)	$ 530,000

Fuel Oil

Steam for leach	$6.4x10^6$BTU/h
Steam for crystalliztion	$10.3x10^6$BTU/h
Building heat	$1.0x10^6$BTU/h
Heat for dryer	$12.3x10^6$BTU/h
Total Heat	$30.0x10^6$BTU/h

Fuel oil, process cost, per year	$1,815,000
Fuel oil, mobile equipment per year	30,000
Total Fuel Oil Cost, per year	$1,845,000

Operating Supplies

Cost as for Section 4.1, Appendix I

Total Operating Supplies Cost	per year	$ 200,000

Maintenance

Maintenance cost (labour and materals) are 3.5% of the total plant capital cost estimate of $33,352,000

Total Maintenance Cost	per year	$1,167,000

Summation

	Per Year
Total Operating Cost	$14,168,000
Contingency, 10% of Total Operating Cost	1,417,000
Total Plant Operating Cost Estimate	$15,585,000
Cost per Kg Mn contained in Product	$1.56

APPENDIX IX

STRIPPED HYDROMETALLURGICAL PROCESS FOR $MnSO_4$

PROCESS AND PLANT DESCRIPTION

The process is identical to that described in Appendix VIII, but with some equipment and circuits deleted from the process flowsheet. The deleted equipment and circuits might not be absolutely necessary for production of manganese sulphate, but the resultant product may not be of salable quality. In addition, these deletions might increase the costs of the remaining circuits and increase operating cost because of a loss of recovery necessitating the treatment of additional ore tonnage. The deletion of spare equipment might also cause cost increases because of a loss of plant availability. These possibilities have not been accounted for in the cost estimates.

The process alternative was considered principally to provide cost variance information for the overall economic review.

DESIGN CRITERIA

All design criteria (basic and engineering) are as for Section 4.8, Appendix VIII.

MATERIAL BALANCE DETAILS

All calculations were based on the material balance for Section 4.8, Appendix VIII.

CAPITAL COST DETAILS

All capital costs are as for Section 4.8, Appendix VIII the conventional hydrometallurgical process for $MnSO_4$ less deletions.

Crushing and Grinding

Deletions:	rock breaker	$ 64,000
	bar grizzley	11,000
	Total	$ 75,000

Total Area Cost = $5,659,000 - $75,000 x 3.5 $ 5,396,000

Neutral Thickening and Filtration

Deletions: none

Total Area Cost $ 1,367,000

Leaching

Deletions:	SO_2 scrubber system	$ 14,000
	scrubber pumps	4,000
	Total	$ 18,000

Total Area Cost = $3,042,000 - $18,000 x 3.5 $ 2,979,000

Acid Filtration

Deletion:	secondary filter stage	
	total equipment	$ 605,000
Total Area Cost = $3,749,000 - $605,000 x 3.5		$ 1,641,000

Impurity Precipitation

Deletion:	total circuit	
Total Area Cost		$0

Manganese Sulphate Crystallization and Product Drying

Deletions:	spare primary exchanges	$ 37,000
	spare secondary exchanger	5,000
	spare autoclave	203,000
	total	$ 245,000
Total Area Cost = $7,861,000 - $245,000 x 3.5		$ 7,003,000

Tailing Neutralization

Deletions:	none	
Total Area Cost		$ 452,000

Reagent Storage and Distribution

Deletions:	none	
Total Area Cost		$ 1,358,000

Process Air, Vacuum and Steam

Deletions:	spare compressor	$ 60,000
	spare vacuum pump	27,000
	spare boiler	65,000
	Total	$ 152,000

Total Area Cost = $1,386,000 - 152,000 x 3.5	$ 854,000

Summation

Total Plant Cost	$21,045,000
Mobile Equipment	544,000
Contingency, 20% of Total Plant Cost	4,209,000
Total Plant Capital Cost Estimate	$25,798,000

OPERATING COST DETAILS

All operating costs are as for Section 4.8, Appendix VIII, the conventional hydrometallurgical process for $MnSO_4$ less deletions.

	Per Year
Reagents	
Deletions: 11 tonne/year flocculant	$ 34,000
Total Reagent Cost = $8,197,000 - $34,000	$ 8,163,000
Labour	
Deletion: one operator per shift - 4 shifts	$ 161,000
Total Labour Cost = $2,228,000 - $161,000	$ 2,067,000
Electrical Energy	
Deletion: 1.2 x 10^6 kWh/year	$ 60,000
Total Electrical Energy Cost = $530,000 - $60,000	$ 470,000
Fuel Oil	
Deletion: none Heat required: 30.0 x 10^6 BTU/h	
Total Fuel Oil Cost	$ 1,845,000

Operating Supplies

Deletion: 35 filter cloths at $600 each	$ 21,000
Total Operating Supplies Cost = $200,000 - $21,000	$ 179,000

Maintenance

Maintenance cost is 3.5% of the total plant capital cost estimate of $25,798,000.

Total Maintenance Cost	$ 903,000

Summation

	Per Year
Total Operating Cost	$13,628,000
Contingency, 10% of Total Operating Cost	1,363,000
Total Plant Operating Cost Estimate	$14,991,000
Cost per kg Mn contained in Product	$1.50

APPENDIX X

PRE-CONCENTRATION AND VAT LEACHING FOR THE CONVENTIONAL HYDROMETALLURGICAL PROCESS FOR $MnSO_4$

PROCESS AND PLANT DESCRIPTION

This process includes the unit operations described for the processes in Section 4.5 and 4.8 (Appendix V and VIII).

Run-of-mine ore is crushed and then ground in a SAG mill. The coarsely ground product is concentrated in spiral concentrators with the concentrate being pumped into prepared leaching vats. The concentrator rejects are pumped to the tailing basin.

When a vat is filled with concentrate the grinding solution is drained and leaching commences with prepared leach solution composed of recycle mother liquor, sulphuric acid and sulphur dioxide. The vats are leached in tandem to maintain pregnant solution grade.

A bleed of pregnant solution or primary vat leach solution is pumped to the impurity precipitation circuit where the pH is adjusted with lime to precipitate iron and other impurities. The precipitated solids are discharged to tailings neutralization.

The purified pregnant solution is preheated with flashed vapour and then brought up to crystallization temperature by heating with steam in heat exchangers and autoclaves to 170°C. Manganese sulphate, $MnSO_4.H_2O$ crystallizes out of the solution. The slurry is cooled by flashing and the crystals are recovered in a thickener-filter circuit. The crystals are dried and dehydrated in a directly fired multi-hearth furnace. The recovered mother liquor is recycled to the leaching vats.

DESIGN CRITERIA

The basic design criteria for this process alternative for $MnSO_4$ production are:

Ore Tonnage	- annual	401,077 tonne/year
	- average	50.64 tonne/h
	- design	55.71 tonne/h
Ore Grade		7.57% Mn
Preconcentration Design	- feed	55.71 tonne/h
	- concentrate	15.31 tonne/h
	- tailings	40.40 tonne/h
	- recovery, manganese	52.28%
	weight	27.48%
Recovery from Concentrate		
	- leaching	65.0%
	- overall	63.0%
Lead Solution Bleed to Process		18.6 tonne/h
Acid Consumption	- concentrate	180 kg/tonne
	- ore basis	49.4 kg/tonne ore
SO_2 consumption	- concentrate	90 kg/tonne
	- ore basis	24.7 kg/tonne ore
Lime consumption	- concentrate	81 kg/tonne
	- ore basis	22.3 kg/tonne ore

The engineering design criteria for equipment sizing and cost estimating are as follows:

Crushing and Grinding

Design criteria as for Section 4.5, Appendix V.

Spiral Concentration

Design criteria as for Section 4.5, Appendix V.

Vat Leaching

Mother Liquor Recycle to Leach 8.5 tonne/h

Mother design criteria as for Section 4.5, Appendix V.

Solution Handling

Design criteria as for Section 4.5, Appendix V.

Impurity Precipitation

Design criteria as for Section 4.1, Appendix I.

Manganese Sulphate Crystallization and Product Drying

Design criteria as for Section 4.8, Appendix VIII.

Tailing Neutralization

Design criteria as for Section 4.1, Appendix I.

Reagent Storage and Distribution

Design criteria as for Section 4.1, Appendix I.

Process Air, Vacuum and Steam

Design criteria as for Section 4.1, Appendix I.

MATERIAL BALANCE DETAILS

A material balance suitable for equipment size and reagent consumption calculation was determined.

The process flows determined were:

	Solids tonne/h	Solution tonne/h
Crushing plant feed	120	-
SAG mill feed	55.71	-
Concentrate production	15.31	-
Vat leach solution flow	-	360
Pregnant solution flow to purification	-	18.6
Impurity solids precipitated	5.8	
Pregnant solution to crystallization	-	19.5
Vapour loss	-	2.5
Mother liquor recycle	-	8.5

CAPITAL COST DETAILS

Crushing and Grinding

Cost as for Section 4.5, Appendix V.

Total Area Cost $ 6,188,000

Spiral Concentration

Cost as for Section 4.5, Appendix V.

Total Area Cost $ 536,000

Vat Leaching

Cost as for Section 4.5, Appendix V.

Total Area Cost $ 486,000

Solution Handling

Cost as for Section 4.5, Appendix V.

Total Area Cost $ 527,000

Impurity Precipitation

Costs for the solids handling portion and solution handling portion are factored using the 6/10 Rule and the cost of Section 4.1, Appendix I.

Solids handling portion:

Equipment cost = $\$329,000 \left(\frac{5.8}{9.73}\right)^{0.6}$ $ 241,000

Solution handling portion:

Equipment cost = $\$505{,}000 \left(\frac{19.4}{45.56}\right)^{0.6}$	\$ 303,000
Total equipment cost	\$ 544,000
Total Area Cost = \$544,000 x 3.5	\$ 1,904,000

Manganese Sulphate Crystallization and Product Drying

The costs for the solids handling portion of the circuit are the same as for Section 4.8, Appendix VIII and the costs for the solution handling portion of the circuit are factored (6/10 Rule) using the appropriate costs of the same Section.

Solids handling portion:

Equipment cost	\$ 848,000

Solution handling portion:

Equipment cost = $\$1{,}398{,}000 \times \left(\frac{19.6}{36.5}\right)^{0.6}$	\$ 960,000
Total equipment cost	\$ 1,808,000
Total Area Cost = \$1,808,000 x 3.5	\$ 6,328,000

Tailing Neutralization

Costs are factored as the basis of solids flow and cost for Section 4.1, Appendix I using the 6/10 Rule.

Total Area Cost - $\$452{,}000 \left(\frac{5.8}{30.58}\right)^{0.6}$	\$ 167,000

Reagent Storage and Distribution

Costs as for Section 4.1, Appendix I.

Total Area Cost	\$ 1,358,000

Process Air, Vacuum and Steam

Vacuum requirements as for Section 4.5, Appendix V.

Two 200 hp boilers will provide the needed steam

Vacuum equipment cost	$	24,000
Steam boiler cost	$	111,000
Total equipment cost	$	135,000
Total Area Cost $135,000 x 3.5	$	473,000

Summation

Total Plant Cost	$17,967,000
Front end loader to feed crusher	511,000
Fork lift	33,000
Contingency, 20% of Total Plant Cost	3,593,000
Total Plant Capital Cost Estimate	$22,104,000

WD1

OPERATING COST DETAILS

Reagents

Reagent consumptions were calculated on the basis of the assumptions for Sections 4.5 and 4.8 (Appendices V and VIII) and the circuit solution balance.

	Consumption kg/tonne conc	Annual Consump, tonne	Unit Cost $/tonne	Total Cost $
H_2SO_4	180	19,840	130	2,579,000
SO_2	90	9,920	200	1,984,000
CaO	81	8,930	98	875,000
Flocculant	0.1	11	3100	33,000
Total Reagent Cost/year				5,471,000

Labour

Cost as for Section 4.5, Appendix V.

Total Labour Cost/year $1,992,000

Electrical Energy

	kW
Crushing and grinding	912.80
Spiral concentration	65.00
Solution handling	148.50
Impurity precipitation	86.00
Manganese sulphate crystallization and product drying	89.00
Tailing neutralization	11.00
Reagent storage and distribution	24.70
Process air, vacuum and steam	45.00
Lighting allowance	120.00
Total	1502.00

Annual electrical energy (assume 80% usage) 9.5×10^6 kWh

Total Electrical Energy Cost/year	$ 476,000

Fuel Oil

Product crystallization	5.5×10^6 BTU/h
Building heat	1.0×10^6 BTU/h
Product drying	12.3×10^6 BTU/h
Total heat	18.8×10^6 BTU/h

Fuel oil, process heat/year	$ 1,138,000
Fuel oil, mobile equipment/year	30,000
Total Fuel Oil Cost/year	$ 1,168,000

Operating Supplies

Cost as for Section 4.5, Appendix V.

Total Operating Supplies Cost/year $ 305,000

Vat Preparation

Cost as for Section 4.5, Appendix V.

Total Vat Preparation Cost/year $ 423,000

Maintenance

Maintenance costs (labour and materials) are 3.5% of the total plant capital cost estimate of $22,104,000.

Total Maintenance Cost/year $ 774,000

Summation

	Per Year
Total Operating Cost	$10,579,000
Contingency, 10% of Total Operating Cost	1,058,000
Total Plant Operating Cost Estimate	$11,637,000
Cost per kg Mn contained in produce	$1.16

APPENDIX XI

REFERENCE PAPERS

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RECENT RESEARCH ON LEACHING MANGANESE

Peter G. Chamberlain

John E. Pahlman

Charles A. Rhoades

U.S. Department of the Interior
Bureau of Mines
Minneapolis, Minnesota

For presentation at the SME-AIME Fall Meeting and Exhibit
Salt Lake City, Utah - October 19-21, 1983

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RECENT RESEARCH ON LEACHING MANGANESE

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Abstract. Domestic mining will not produce a significant portion of the manganese consumed in the United States unless new mining and processing techniques are implemented that will enable domestic lean ores to economically compete with rich foreign ores. One mining-processing technique that appears promising for certain deposits is in situ and/or heap leaching with aqueous SO_2. Since leaching offers economic advantages of low capital costs, low operating costs, and short start-up times as well as environmental and safety improvements over conventional mining-processing, the Bureau of Mines studied the leachability of manganese deposits from several districts. Manganese oxides, hydroxides, and carbonates were leached to 80 to 90 pct recovery in 7 to 40 days of leaching with a solution containing 5 pct SO_2 in column leaching tests. Preliminary cost analyses of surface mining-heap leaching, block caving-in situ leaching, and shrinkage stoping-in situ leaching indicate that the products of manganese leaching may be economically produced for the chemical and electrolytic manganese industries. The easiest marketable product to obtain would be a manganese sulfate commonly used as a fertilizer amendment. Manganese carbonates and hydroxides could also be precipitated from the pregnant leach solutions.

Introduction

Domestic mines account for a mere 2 pct of the manganese consumption in the United States. Although plentiful supplies of manganese are readily available from several foreign countries, the Nation's vulnerability to sudden cessation of foreign shipments plus the current trade imbalance has stimulated a renewed search for economic methods of producing manganese from domestic deposits. The Bureau's Minerals Availability Field Office recently published a study (Kilgore and Thomas, 1982) that manganese could not be produced competitively from 8 domestic deposits using various conventional mining and milling schemes. The long-run total costs for production from these domestic operations ranged from \$8 to \$35/long ton unit or 22.4 lbs (\$.16 to \$.71/kg) contained manganese as compared to 1982 prices averaging between \$1.60 and \$1.70/long ton unit (\$.03/kg) for 48 pct minimum grade manganese ore imported f.o.b. Chicago or Pittsburgh.

Clearly, significant new technology is necessary to enable domestic producers to produce manganese at competitive prices. From strategic material and market size standpoints, such new technology would have the greatest impact if it provided manganese in the form that could be used directly by the steel industry. The chemical, battery, and superalloy industries, however, annually consume about 100,000 tons (10×10^6 kg) of manganese and it is these markets that seem attractive targets for the technology recently researched by the Bureau—in situ and/or heap leaching manganese with aqueous SO_2. Although the products of leaching cannot be charged directly into steelmaking furnaces without expensive reprocessing, they can be readily incorporated into other industries as manganese sulfates, oxides, and carbonates.

In situ leaching has been very successful for uranium and, to a lesser degree, for copper. Heap leaching has become a very important source of gold, silver and copper. Leaching offers advantages of low capital cost, low operating cost, and short start-up times as well as environmental and safety improvements over conventional mining. The primary disadvantage—low recovery—favors consideration of the method for small and/or low grade deposits that would not otherwise be mined. Most domestic manganese deposits fall in that category.

Leaching manganese with SO_2 was explored by the Bureau during the 1940's (Wyman and Ravitz, 1947). Experiments demonstrated that a variety of manganese deposits could be leached with SO_2 to produce manganese sulfate solutions. Laboratory leaching tests led to pilot plant experiments and eventually to commercial production of manganese from a vat leaching operation during World War II. Although heap leaching experiments were conducted in the 1950's, (Bender and Rampacek, 1957), in situ leaching was not seriously considered until Mr. James Lake presented several ideas on leaching the Artillery Peak deposits to the Bureau. A contract subsequently awarded to a private firm to evaluate feasibility of in situ leaching various metals concluded that in situ leaching of massive oxide manganese deposits offered potential for commercialization for specialized manganese markets (chemical products) (Potter, et al., 1982; Potter, Chase, Chamberlain, 1982).

The favorable consideration given to leaching manganese triggered an in-depth evaluation of leaching potential, summarized in this paper. The approach was basically to: 1) rank deposits based on geologic, hydrologic, and geographic factors; 2) collect samples from a variety of the types of deposits that were highly ranked; 3) conduct laboratory leaching tests on the samples with aqueous SO_2; 4) analyze technical feasibility and economics of leaching several typical types of deposits; and 5) conduct field tests to verify laboratory results and certain assumptions in the economic analysis. Budgetary constraints have, however, prevented any field testing to date.

Early in the investigations, a special circumstance of leaching manganese presented interesting possibilities—leaching silver ores that are refractory to ordinary cyanidation because of their manganese content. A dual leaching system consisting of leaching the manganese out of the ore with aqueous SO_2 followed by leaching the silver with cyanide (after flushing with NaOH or lime to remove any acid) appeared promising for greatly increasing silver recovery as well as providing salable manganese. Results of experiments with such a method are presented separately (Rhodes, Pahlman, Chamberlain, 1984).

Evaluation of In Situ Leaching for Manganese

Background data were gathered on as many manganese deposits as possible. To provide an objective rationale for selecting deposits to be used in the laboratory testing program and to establish "typical" conditions as a basis for conceptual designs and cost analysis, an arbitrary ranking process was implemented. Ranking factors included geologic, hydrologic, and geographical considerations. A variety of highly ranked deposits were selected for the testing program and subsequent evaluation of leaching potential. An attempt was made to ensure representation of different manganese minerals (oxides, hydroxides, carbonates, silicates), physical characteristics, geologic origins, and geographic locations in the selected list.

Characteristics of Selected Deposits

The deposits that were sampled and tested in column leaching experiments are characterized in table 1. The higher order oxide minerals predominate—pyrolusite, psilomelane, and wad (an amorphous assemblage of oxides). Hydrothermal vein deposits are the most numerous although the sedimentary deposits are larger and may represent more viable leaching targets. The sedimentary deposits are generally massive, low grade, and high volume compared to the higher grade, small volume, vein deposits. With the exception of the Cuyuna deposits, the sedimentary beds were horizontal while most of the vein deposits are steeply dipped.

Laboratory Tests

Although many laboratory tests have been conducted on leaching manganese ores with aqueous SO_2 (summarized in Chamberlain, 1981), the testing program established at the TCRC had several key differences from those earlier tests. For evaluating heap and/or in situ leaching possibilities, relatively large size ore pieces (compared to the more common 100 mesh size) were tested since they are the ones that control leaching rate, recovery, and profitability. Percolating leach solutions through ore columns rather than immersing samples in beakers was the preferred testing method because it more closely simulated operating conditions. A final difference from earlier programs was the use of low concentration of SO_2 in solution. Tests conducted on sea nodules (Khalafalla and Pahlman, 1981) earlier had demonstrated that at low concentrations, SO_2 was more selective in the metals it leached than it was at high concentrations. At a 5 pct solution, most of the MANGANESE would be leached but very little of the iron and other metals would be co-leached.

Column leaching tests were conducted in 4 in (10 cm) columns that were 3 ft (3m) long and contained 3.5 kg of -1 in +1/2 in (2.3 x 1.3 cm) pieces. A 5 pct SO_2 solution was dripped into the column at 1 ml per minute. This rate was fairly close to the application rates in gold/silver and copper heap leaching operations. At that rate the entire column remained unsaturated. The tests were usually run until the manganese in the effluent from the column dropped to about 1 g/L (figure 1). During any one test at least 50 L of solution was applied except for the columns of Artillery Peak ore in which 25 to 30 L of solution extracted most of the manganese.

Manganese oxides, carbonates, and even silicates were all leachable in column leaching tests (figure 2, table 2). Although the leaching rate was highest for Artillery Peak samples, the higher grade of materials from other sites may provide effluent solution grade equal to or greater than that obtained from Artillery Peak. Greater than 80 pct extraction was obtained for most manganese oxide ores. Although batch tests on finer particles of ore conducted in flasks had earlier indicated minimal manganese leaching from carbonates and silicates, the percolating action of solutions in column leaching yielded recoveries from carbonate samples of greater than 90 pct. Leaching of manganese silicates, although slow, resulted in solubilization of 1 to 2 g of manganese per liter of leach solution applied.

Several generalizations can be stated about leaching manganese ores in the columns. First, the leaching rate of most ores tested is not dependent upon the solubilization reaction between the manganese and aqueous SO_2, but instead upon the rate of solution ingress and egress from ore particles. Batch tests on several oxide samples demonstrated that, if ground to minus 60 mesh, more than 90 pct of the manganese was leached within a few minutes compared to the several days required of the large pieces used in the column leaching tests.

Second, the leaching rate of manganese is affected by gangue minerals present in the ore. Calcium for instance, is readily solubilized by SO_2 solutions which diminishes the opportunity for manganese dissolution. This was evidenced by the results of leaching tests on two ores with similar manganese and iron contents, but different calcium contents. After passing 90 L of 5 pct SO_2 solution through the column of each ore, the one with 5.7 pct calcium showed 73 pct manganese extraction while the one with only 1.0 pct calcium demonstrated 95 pct manganese extraction. Also iron as goethite or carbonate in an ore competes with the manganese for the SO_2. Although SO_2 has more affinity for manganese then iron, when the top layers of ore in the column are depleted of manganese, the SO_2 will still pass in and out of these ore pieces leaching the next available metal.

Third, although there was a spectrum of manganese minerals in the samples tested, the majority of samples that contain either pyrolusite, psilomelane, wad, or rhodochrosite yielded manganese recoveries in the 80 to 95 pct range.

It is also interesting to note that rhodonite, a manganese silicate that is supposedly inert to SO_2, appears to be leachable, although at a slow rate. This ore continued to leach at 1 to 2 g of manganese per liter even after application of 70 L of solution to the column.

Evaporation tests were made to identify the manganese species in the leach solution. Research in the 1940's indicated that the manganese in solution would be predominantly as a sulfate with some dithionate, which x-ray analysis of the evaporation residue from the column leaching tests confirmed for deposits that contained high-valence manganese. Manganese in leach solutions obtained from manganese in the plus 2 valence, was primarily a bisulfite which decomposed to sulfite and SO_2 upon heating.

Tests were also conducted on techniques for extracting manganese from the leach solutions. The easiest technique was to simply evaporate the solution to obtain manganese sulfate that is directly marketable as a fertilizer component. By raising the pH of the leach solution to 6 and filtering it prior to evaporation, ferrous and ferric iron were precipitated while manganese remained in solution. As an alternative to evaporation, the filtered solution was reacted with a base to precipitate the manganese as hydroxide that upon drying oxidized to either Mn_3O_4 or MnO_2. This product could be used as feed for further processing by battery and chemical industries. In yet another alternative, the iron-free solution was sparged with CO_2 to precipitate manganese carbonate that, upon heating, decomposes to an oxide. These products are of interest to battery and chemical industries or can be redissolved for electrowinning metallic manganese.

Conceptual Design and Cost Analyses

With the wide variety of manganese deposits in this country that are potentially leachable, an equally wide variety of mining/leaching schemes can be considered. Basically any system that is or would be developed for mining an orebody of certain configuration and physical properties becomes a prime candidate system for preparing a manganese orebody of similar configuration and properties for leaching. For instance, a shallow orebody that would be mined from an open pit for milling would, for a leaching system, be mined from an open pit and heap leached. Deeper deposits could be broken for leaching by shrinkage stoping (figure 3), long-hole open stoping-stoping, sublevel stoping, other overhand stoping, or by block caving methods exactly as they would be mined for conventional milling. The prime difference of a leaching system from a conventional system is that only enough rock would be hauled to the surface to provide room for the remaining rock to expand during the breakage process; the 20 to 25 pct hauled to the surface would be heap leached while the remainder would be leached in place underground. The underground oil shale retorts created by Occidental Oil Shale at their test facility in Colorado also seem reasonable patterns for creating underground chambers in which massive manganese deposits could be leached (figure 4). The primary cost savings with any of these systems becomes obvious: reduced haulage/hoisting requirements and no expensive milling operations.

Leaching systems would employ aqueous SO_2 created by burning pyrite or elemental sulfur in a fluidized bed and bubbling the combustion gases through water (Potter, et al., 1982) (figure 5). This latter step could be done adjacent to the sulfur burner and the aqueous SO_2 pumped to the broken ore or the gas could be piped under the broken ore where it would rise to mix with water percolated downward from the top of the ore pile. Mining consultants estimate that costs for this system would run \$0.01 to \$0.03/lb (\$.005 to \$.014/kg) of manganese produced by leaching.

The effluent from the leaching phase would be a solution containing a mixture of manganese sulfate, dithionate and (for carbonate ores) bisulfite. Effluent could be collected, recharged with SO_2 and recycled to the heap if higher grade solutions were desirable. The easiest marketable product to produce from the effluent leach solution would be manganese sulfate commonly used as a fertilizer amendment or for other purposes by the chemical supply industry and recently selling for \$250/ton (\$.057/kg). The sulfate can be crystalized directly by evaporating the solution.

Another product interesting to the chemical industry, manganese carbonate, can be precipitated by adding $NaCO_3$ to the effluent solution. Alternatively manganese hydroxide can be precipitated by neutralizing the solution with lime and heating the precipitate to form useful manganese oxide. Electrowinning to produce metallic manganese is also possible.

In determining the economic feasibility of mining and leaching domestic manganese deposits, four different mining scenarios were chosen to represent the ore preparation systems that would be suitable for leaching a majority of the domestic deposits: surface mining, block caving, sublevel stoping, and shrinkage stoping. Costs for these systems were estimated using equations from a handbook prepared by STRAAM Engineers under contract with the Bureau of Mines (Clement, et. al., 1977) for feasibility estimates of capital and operating costs for mining non fossil-fuel mineral deposits. Capital costs include roads, utility sources, buildings, town site, drilling equipment, loaders, haulage equipment, SO_2 generation equipment, solution distribution equipment, engineering and construction fees, and 60 day's operating expenses. Operating costs include land clearing, overburden removal, core drilling, drilling and blasting, shaft sinking, ore excavation and hauling, SO_2 generation, solution distribution, evaporation pond maintenance, restoration, general operations costs, water supply, drainage waste disposal, transporation and administrative costs. An operating life of 10 years was assumed. Since reasonable cost estimates require selection of representative geologic, geographic, and physical parameters, two deposits were chosen to provide input data for the cost estimates. One was a low-grade massive deposit —the Artillery Peak deposit; the other was a thin vein deposit typical of many western ores—the Paymaster district.

Details of the economic analysis are provided in a forthcoming Bureau of Mines Report of Investigations. In summary surface mining and heap leaching would be the cheapest method for leaching wherever it would be possible. Costs (based on 1982 dollars) for an exposed massive deposit 100 m thick of 3 to 4 pct manganese would run $219/ton ($.050/kg) for $MnSO_4$ if 60 pct of the manganese was leached in an operation sized to produce 9,300 tons/d (8.4×10^6 kg/d) of $MnSO_4$ for 10 years (table 3). This product has recently sold for $250/ton ($.057/kg) for agricultural fertilizer. Massive deposits typical of the Artillery Peak, Cuyuna, and Aroostook districts plus several of the larger vein deposits would appear amenable to such a system.

Deeper massive deposits that could be mined and leached in block caving system would generate operating costs $274/ton ($.062/kg) for $MnSO_4$ produced, assuming that 60 pct of the manganese in the caved zones could be leached and that the deposit had an ore grade of 3.5 pct manganese. This would dictate that an ore grade of around 5 pct would be necessary to cover all of the operating costs plus pay back the capital costs with reasonable profit. Of the deposits evaluated in this study, selected portions of the Artillery Peak district are the only ones with the physical characteristics that would favor caving and have the required ore grade.

For massive deposits that would not cave readily, a leaching system encorporating sublevel stoping for ore preparation was conceptualized. Operating costs for such a leaching system would approach $320/ton ($.073/kg) $MnSO_4$ for a 3.5 pct manganese grade deposit. Again, to cover all operating costs and repay a profit on the capital expenditures an ore grade of 6 or 7 pct would probably be required to develop an operation. High grade portions of the Artillery Peak district containing "hard" matrix-cementing minerals, the Aroostook, and perhaps several of the larger vein deposits would seem the most likely candidates for such a system.

Narrow vein deposits would be more expensive to mine and leach underground. For instance, our feasibility analysis indicated that if a 10 to 30 ft (3 to 10 m) thick steeply dipping vein containing 10 pct manganese were mined by shrinkage stoping methods (figure 3) and leached, the operating cost would be $335/ton ($.076/kg)1 of $MnSO_4$. The practical application of such a system, therefore, would probably be limited to 15 pct or higher grade deposits. Several such deposits exist in Colorado, Montana, and Nevada.

A unique and particularly interesting scenario could be developed for leaching manganese in the Cuyuna Iron Range of Minnesota. Significant manganese tonnages are contained in vast waste rock piles from the depleted iron mines scattered profusely throughout the district. Many of the piles could be moved short distances to prepared pads and heap leached (with or without crushing). Costs were not developed for such a system. Similarly, coarse tailing in the Leadville, Colorado, and Cuyuna districts contain significant amounts of manganese that could be directly leached without any additional ore preparation costs.

Another unique leaching situation is created by the vast horizontal, unconsolidated deposits of nodules containing manganese centered around Chamberlain, South Dakota. The low grade and impervious nature of the beds has thwarted all attempts to recover the menganese from the high grade nodules. An intriguing possibility would be to "leach farm" the beds, 1 ft (.3m) at a time. In leach farming (Lankenau and Lake, 1973) the top layer would be worked with rippers and farm machinery to mix the leach solution with the shale matrix containing nodules for a depth of 1 ft (0.3 m). After a "soak" period the layer would be removed by scraper or bulldozer and washed into a settling pond. The pregnant leach solution would then be drawn out of the pond for extracting the manganese. No costs for this conjectural leaching system were developed.

Costs were not developed for any leaching product other than $MnSO_4$. Although the technology is readily available, it is likely that precipitating manganese oxides or carbonates from the

Cooling crystallization.
700 gpl at 70°C
520 gpl at 5°C

original $MnSO_4$ solution would be more expensive than solar evaporation to obtain $MnSO_4$.

Likewise, $MnSO_4$ could be used as the raw material for electrowinning metallic manganese but determining the costs of such a process were beyond the scope of this study. One refinery purchases 74 pct MnO_2 for $130/ton ($.034 kg) but since that material must be roasted before processing (while the $MnSO_4$ from a leaching operation does not) and contain undesirable contaminants, a direct cost comparison cannot be made.

Conclusions

The heavy reliance of this country on foreign imports of manganese has stimulated the search for new and innovative methods of producing manganese from small and/or low grade domestic deposits. One method that appears to offer hope for fulfuilling some non-ferromanganese requirements is in situ/heap leaching. Laboratory column leaching tests on 19 samples from 13 mining districts indicate that manganese oxides, hydroxides and carbonates were leached to 62 to 95 pct extraction with aqueous SO_2. The leaching product is a solution containing $MnSO_4$ that can be precipitated by evaporation for direct marketing as an amendment for agricultural fertilizer. Alternatively manganese oxide or carbonate can be precipitated from the solution for sale to the chemical industry, battery manufacturers, or possibly to processing plants that produce metallic manganese by electrowinning.

A very preliminary cost feasibility analysis based on 60 pct recovery of manganese as $MnSO_4$ valued at $250/ton ($.057/kg) from different mining/in situ leaching scenarios for oxide and carbonate deposits indicated that the following strategies were potentially viable: (1) surface mining of massive deposits containing more than 4 pct manganese, (2) block caving of massive deposits containing more than 5 pct manganese, (3) sublevel stoping of massive deposits containing more than 7 pct manganese and (4) shrinkage stoping of thin vein deposits containing more than 15 pct manganese. Although no cost analysis were developed for them, occurrences of manganese in iron mine waste rock piles in the Cuyuna range and in coarse tailings from several districts seem to offer particularly interesting possibilities for heap leaching with no ore preparation.

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TABLE 1. - Manganese sample characteristics

State	District	Manganese form	Primary manganese minerals	Grade[1] (pct)	Origin
Arizona	Artillery Peak	Oxide	Wad[1]	4	Sedimentary
Do	Tombstone	Do	Psilomelane Manganite Pyrolusite	2	Hydrothermal vein
Do	Do	Do	Manganite Psilomelane Pyrolusite	10	Do
Do	Harshaw	Do	Cryptomelane Psilomelane Pyrolusite	28	Do
Do	Do	Do	Pyrolusite	11	Do
Arkansas	Batesville	Do	Hausmannite Psilomelane Braunite	7	Do
California	Paymaster	Do	Pyrolusite Psilomelane Manganite	20	Do
Colorado	San Juan Mtn.	Silicate	Rhodonite[2]	30	Do
Do	Leadville	Oxide	Psilomelane Pyrolusite	11	Do
Maine	North Aroostook	Oxide/Silicate	Braunite Rhodochrosite	23	Submarine volcanogenic-sedimentary
Minnesota	Cuyuna	Oxide/Hydroxide	Manganite	3	Sedimentary
Do	Do	Carbonate	Rhodochrosite	8	Do
Do	Do	Oxide	Pyrolusite Stilpnomelane Manganite	6	Do
Do	Do	Do	Pyrolusite	11	Do
Do	Do	Do	Pyrolusite	4	Do
Nevada	Virgin River	Do	Wad	27	Do
Do	Boulder City	Do	Wad	9	Do
Do	Three Kids	Do	Wad	16	Do
South Dakota	Chamberlain	Carbonate	Manganocalcite Pyrolusite Manganite	[3]1	Fossil sea nodules

NA = Not available.

[1] Amorphous collection of oxides.

[2] May be the pyroxmanganite variety.

[3] Grade of manganese in shale. Nodules average 18 to 20 pct manganese.

TABLE 2. - Manganese recovery from SO_2 leaching

District	Manganese form	Grade, pct Mn	Manganese recovery pct[1]
Artillery Peak	Oxide	4	95
Tombstone	Do	2	82
Tombstone	Do	10	92
Harshaw	Do	28	95
Harshaw	Do	11	78
Batesville	Do	7	59
Paymaster	Do	20	67
San Juan Mtn.	Silicate	30	30
Leadville	Oxide	11	83
North Aroostook	Oxide/Silicate	23	93
Cuyuna	Oxide/Hydroxide	3	92
Cuyuna	Carbonate	8	94
Cuyuna	Oxide	6	61
Cuyuna	Do	11	62
Cuyuna	Do	4	92
Virgin River	Do	27	73
Boulder City	Do	9	82
Three Kids	Do	16	94
Chamberlain	Carbonate	18	89

[1] Generally based on results of 1-3 tests each.

TABLE 3. - Cost analysis (1982 dollars) of leaching systems

System	$MnSO_4$ capacity, tons/day	Costs		Operating costs per ton $MnSO_4$
		Capital, millions	Daily operating	
Surface mine, heap leach	9,300	14.1	$66,500	[1]$219
Sublevel stoping	9,300	NA	89,700	[1] 321
Block caving	9,300	16.2	80,700	[1] 274
Shrinkage stoping	800	4.9	26,800	[2] 335

NA = Not analyzed.

[1] Massive deposit containing 3.5 pct manganese.

[2] Vein deposit contains 10 pct manganese.

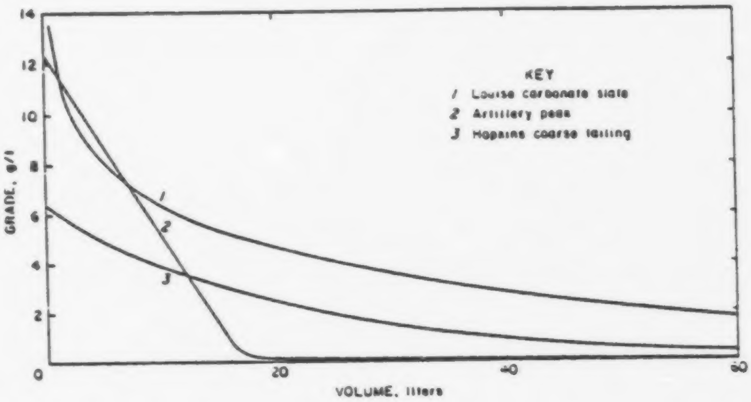

Figure 1. – Concentration of manganese in effluent leaching solution

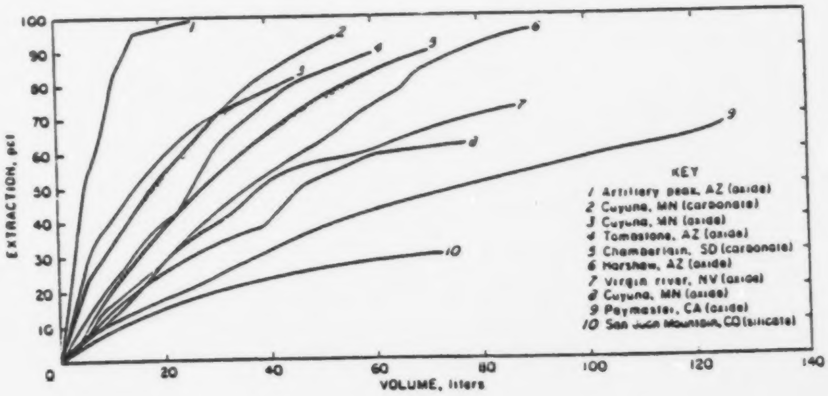

Figure 2. - Selected leaching recovery curves for samples from several manganese sources

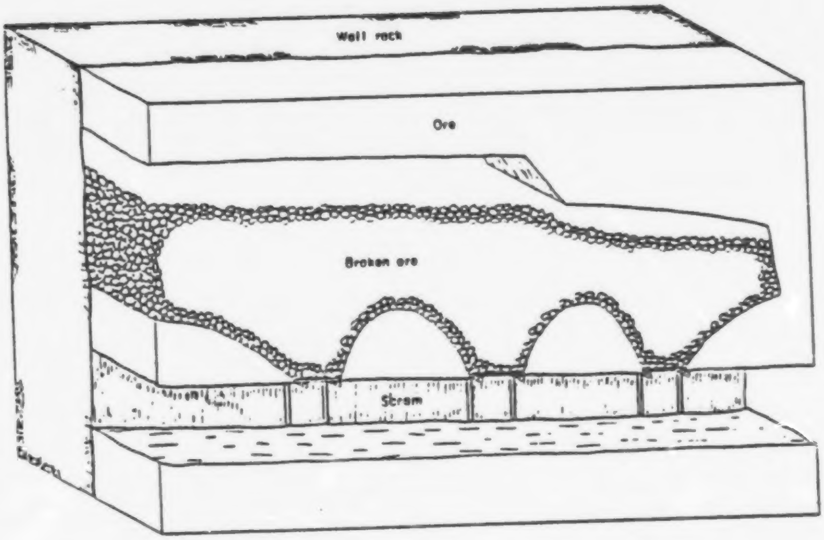
Wall rock
Ore
Broken ore
Stream

Figure 3a. - Ore preparation phase, shrinkage-stope mining/in situ leaching

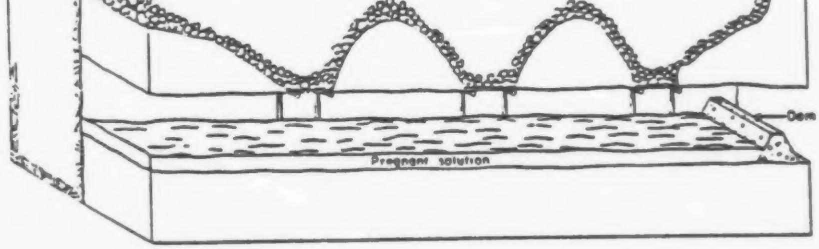
Solution injection wells
Ore
Broken ore
Dam
Pregnant solution

Figure 3b. - Leaching phase, shrinkage-stope mining/in situ leaching

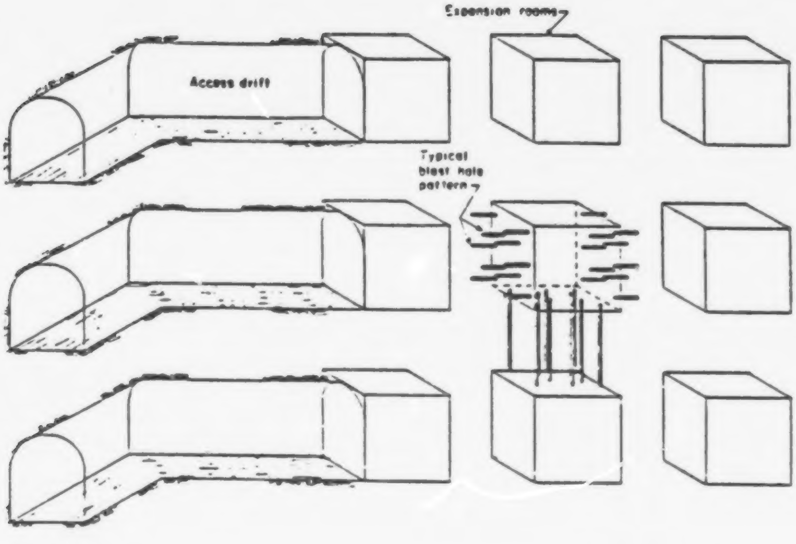
Expansion rooms
Access drift
Typical blast hole pattern

Figure 4a. - Blasting scheme (cross-section), underground leaching concept patterned after modified in situ retort for oil shale

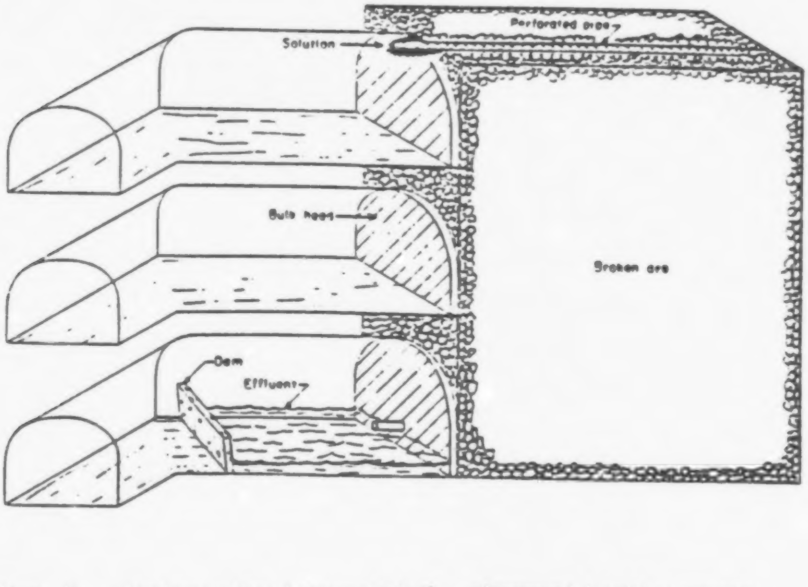
Perforated pipe
Solution
Bulk head
Broken ore
Dam
Effluent

Figure 4b. - Leaching phase (cross-section), underground leaching concept patterned after modified in situ retort for oil shale

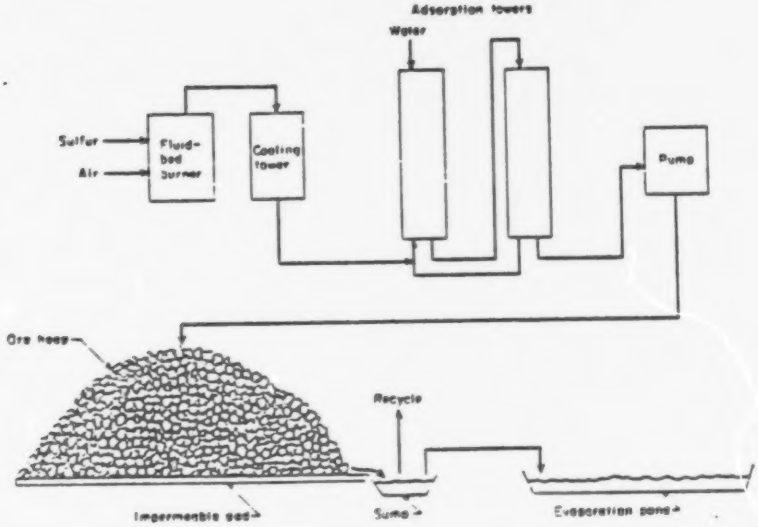
Adsorption towers
Water
Sulfur
Air
Fluid-bed burner
Cooling tower
Pump
Ore heap
Recycle
Impermeable pad
Sump
Evaporation pond

Figure 5. - Leaching system for manganese (schematic)

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Arizona's Artillery Peak Manganese Deposits and Their Potential for Leaching

Abstract. The Bureau of Mines is conducting research to boost domestic production of manganese. Virtually no mining exists in the low grade and/or small deposits that occur in this Country. One large, low-grade manganese deposit located in westcentral Arizona--The Artillery Peak district--was extensively studied and sporadically mined during the 1940's and 1950's. Although manganese mining in the district has been long dormant, recent studies indicate that the deposits may be amenable to economical production by heap and/or in situ leaching methods. A conceptual design for an in situ leaching system features undercutting and blasting a block of ore, leaching with SO_2, and extracting the manganese as a sulfate or oxide. Possible applications to other manganese deposits depend upon formation permeability and specific manganese mineralogy.

Introduction

Manganese is a strategic commodity because of its importance in steel production with no recognized substitutes. For many years the United States has almost totally relied upon foreign sources of manganese (Figure 1), importing 98 percent of the annual consumption (Bureau of Mines, 1979). Since much of this manganese originates from politically unstable or possibly unfriendly countries, it is in the National interest to develop a modest annual domestic production.

The dearth of production is not caused by an absence of deposits. Although the United States has no identified reserves, resources abound. While no high-grade deposits (more than 35 percent manganese) are mined, ores ranging 10 to 35 percent manganese were shipped from mines in New Mexico, South Carolina, and Minnesota during the past few years, and during both World Wars manganese was mined from numerous deposits.

If deposits are plentiful in this country, why is there no production from them? The prime cause is the relatively low grade of these deposits compared to those of other countries. The higher grade of the foreign ore coupled with lower labor costs provide the foreign sources with a currently insurmountable economic advantage over their domestic counterparts.

Although there is sentiment for exhausting the foreign sources and hoarding our own, a safer philosophy would seem to favor boosting domestic production capacity to a level that would avert the catastrophic consequences of drastic import disruption. It will be difficult, however, to stimulate any production without new and economical technology.

The purpose of this presentation/publication is to: (1) show that there are ample deposits that may be mined; (2) describe in detail one such deposit that offers opportunity for future production; and (3) describe a relatively new scheme for mining and producing manganese--in situ leaching.

Manganese Deposits in the United States

Manganese deposits were formed when manganese, originally dissolved out of crystalline rocks, was deposited as carbonates, oxides, hydroxides, and silicates. Most of the mined deposits have been oxides or mixtures of oxides that occur as massive deposits in sedimentary or volcanic rocks or as vein filling material.

United States resources contain an estimated 74 million tons of manganese (Bureau of Mines, 1978), mostly in deposits with a grade below 10 percent. A Bureau report on manganese (Dean, DeVaney, and Coghill, 1934), mentions that manganese deposits were found in 300 districts in 35 states. Twelve of these were ultimately identified as possibly desirable for mining and hydrometallurgical processing (Dean, Leaver, and Joseph, 1934) whereas others showed potential for conventional treatment but not at peacetime manganese prices.

The National Materials Advisory Board, in a recent evaluation of the manganese mining industry (Manganese Recovery Technology Panel, 1976), identified several deposits that were the most likely candidates for commercial exploitation (Table 1). Four of these appeared on the previously mentioned list of twelve with hydrometallurgical potential (including the Artillery Peak deposits).

The Artillery Peak district deposits that have attracted much Bureau research in the past and hosted commercial open-pit mining operations 25 years ago once again emerge as a possible proving ground for new technology. Since the deposits illustrate the type that may be amenable to hydrometallurgical techniques, they will be discussed in more detail.

Artillery Peak District

The Artillery Mountains lie in westcentral Arizona, 30 miles (48 km) east of the Colorado River and midway between Wenden and Wikieup (Figure 2). The best deposits generally occur southwest of Artillery Peak in the foothills of the Artillery Mountains north of the Bill Williams River.

The region is characterized by a hot, dry climate and scanty vegetation typical of Arizona's desert regions.

Geology

The geology of the district was extensively studied and reported by the United States Geological Survey in the late 1930's and 1940's (Lasky and Webber, 1944; Ibid, 1949). The manganese is contained in thick alluvial fan and playa sediments (called the Chapin Wash formation) deposited in a large fault basin during early Pliocene age. Alternating layers of basalt flows and sediment subsequently covered the deposits during late Pliocene and Quarternary times. The deposits are somewhat folded and have been considerably faulted (Figure 3).

The manganiferous beds occur in two zones, each 300 to 400 feet (90 to 120 m) thick. The upper zone--the main zone of interest--contains manganese in three types of deposits: sandstone, clay, and a "hard" material. In the sandstone and clay deposits, the manganese appears as amorphous manganese oxides (wad) coating the matrix grains. The "hard" ore is basically a sandstone that was infused with opal and calcite while the original manganese oxides were altered to psilomelane [4 $MnO_2(Mn,Ba,K)O \cdot H_2O$] and manganite ($Mn_2O_3 \cdot H_2O$).

The manganese content of the ore averages 3.5 to 4 percent while iron is about 3 percent. Although the grade changes drastically from bed to bed, the manganese and iron content are quite constant over large volumes of ore. The district is estimated to contain 175 million short tons (1.6×10^{11} kg) of material at about 4 percent. About 20 million tons (1.8×10^{10} kg) contains more than 5 percent and 2 million tons (1.8×10^{9} kg) contains more than 10 percent manganese. Most of the higher grade material is the "hard" ore that is more difficult to process.

History

Sporadic mining activity began on claims in the district during World War I but halted with the conclusion of the War before any ore could be shipped. In the late 1930's the M. A. Hanna Company obtained leases on most of the promising ground and conducted extensive exploration. The USGS and the Bureau of Mines also conducted exploration activities there in the late 1930's and early 1940's as part of the war minerals investigations.

In 1941 the Bureau drove an adit into a high grade zone from the Maggie Canyon and mined several rooms to assess room-and-pillar mining techniques. The rooms remained stable without timbering to widths and heights over 20 feet (Sanford and Stewart, 1948).

About 2,500 tons (2.3 million kg) of ore from the Maggie adit were shipped to the Bureau's experimental plant at Boulder City, Nevada, for metallurgical testing. These tests concluded that flotation was not effective but that hydrometallurgical treatment offered possibilities; high recoveries were obtained by leaching with sulfuric acid and with sulfur dioxide.

In 1942, the Bureau tested sulfur dioxide leaching on manganese samples from 58 deposits including the Artillery Peak district (Wyman and Ravitz, 1947). The Artillery Peak ore leached readily providing a recovery of over 95 percent of the manganese with only small amounts of iron and phosphorus. Artillery Peak ore was subsequently used in many experiments to test variations of sulfur dioxide leaching; the principal techniques were summarized by Henn, Kirby and Norman (1968).

The Russian cessation of manganese shipments in 1949 and the Korean conflict revitalized interest in the Artillery Peak district. A Government Purchasing Depot at Wenden, Arizona received 25,600 tons (23.2 million kg) of ore in 1953 and 55,700 tons (50.5 million kg) in 1954 from the Artillery Peak district of which 9,000 tons (8 million kg) was from the Bureau's experimental mine.

Metallurgical research on the ore at the Bureau's Boulder City laboratory led to a combination flotation-leaching process (Rosenbaum, Schack, Lang, and Clemmer, 1957) that could provide the desired grade manganese but the high silica content disqualified the concentrate as standard feed for ferromanganese production. Concurrent research on a dithionate and sulfur dioxide vat leaching process with lime precipitation of manganese hydroxide was successfully tested at a 1,500 pound/day (680 kg/day) pilot plant scale between 1951 and 1953 (Rampacek, Fuller, and Clemmer, 1959). Smelting tests showed that ferromanganese could be produced from concentrates and blends on concentrates and dithionate leach products (Edlund and Lang, 1962; Petermann and Lang, 1962).

Since the mid 1950's there has been no manganese mining activity although other commodities are mined in the region.

In Situ Leaching Applications for Manganese

The Bureau of Mines devotes a portion of its annual budget to research on in situ or in place leaching. Basically in situ leaching is the dissolution of mineral values from ore left in place and the recovery of metals from the pregnant leach solution. Benefits derived from in situ leaching include: low front-end costs, fast cash flow, reduced health/safety risks, and less environmental impact. The major disadvantage--lower recovery--has channeled industry interest towards small and/or low grade deposits that cannot be profitably mined and milled by conventional methods. This condition plus the above benefits favor consideration of the method for manganese.

Although hydrometallurgical techniques have been considered for many years (for instance, Davis, 1930) and leaching tests with sulfur dioxide in the early 1940's first demonstrated that the method preferentially leached manganese before iron and phosphorus, in situ leaching has not been seriously considered for manganese. Problems with slime build-up in the dithionate-sulfur dioxide vat leaching process developed at the Boulder City facility led to a series of experiments on several types of ore (including two from the Artillery Mountains) by forcing SO_2 up through ore heaps and then percolating downward a wash consisting of either water or a solution containing calcium dithionate (Bender and Rampacek, 1957). Although the experiments proved that manganese could be heap leached, no commercial activity resulted. Meeting with Bureau officials in 1978, Mr. James Lake first outlined ideas for in situ leaching manganese. This prompted an evaluation of leaching feasibility in a contract awarded to Mountain States Research and Development. Manganese received strong attention because it had been identified as a critical commodity without domestic production and it occurred in low grade and/or small orebodies that would not otherwise be mined. After considering potential leaching characteristics of a variety of manganese minerals found in deposits from Maine, Minnesota, and southwestern United States, the Contractor concluded that leaching manganese oxide ores with sulfur dioxide offered the best prospects for commerical development (Potter, Nordhausen, Fletcher, Lake, Rouse and Wojtasiak, 1981).

A conceptual design for leaching manganese developed by Mountain States (Potter, Chase, Chamberlain, 1981, and Potter, Nordhausen, Fletcher, Lake, Rouse and Wojtasiak, 1981) (Figure 4) suggests that massive orebodies could be undercut and broken for leaching in place (Figure 5). Sulfurous acid could be cheaply produced at the site by burning sulfur and bubbling the combustion gases containing SO_2 through water. The acid could then be injected at the top of the broken ore and allowed to percolate downward. Alternatively, the gases could be piped to the base of the broken ore and allowed to bubble upward while water or a recycled solution injected at the top of the ore percolates downward. Recovery of the manganese salts (sulfates and dithionates) could be cheaply archieved in solar evaporation ponds in arid portions of the Country.

Mountain States analyzed such a leaching system based on the deposits in the Artillery Peak district. The reasons for selecting this deposit to represent a "typical" oxide deposit were: (1) Much information is available on the geology. (2) Background data are available from extensive Bureau research conducted in the 1940's and 1950's. (3) The district contains a large resource of low-grade ore. (4) Industry is interested in possible mine development. The economic analysis, based on a final product containing manganese sulfate worth $200/ton ($0.22/kg) as an agricultural chemical, yielded a discounted-cash-flow return-on-investment of 28 percent with a 3-year payout. Such numbers offer sufficient incentive for research to prove or refute their validity and, if valid, to encourage development of leaching operations.

For the strategic ferromanganese market, additional studies are needed on the costs of producing oxides from the manganese sulfate solution. Clues may be obtained from early Bureau research (Prasky, Joyce, and Swanson, 1963) on low temperature decomposition of $MnSO_4$ to the oxide form with liberation of SO_2 (which could then be recycled to the leaching solution). Problems caused by the dithionate salts in the leach product, however, would have to be overcome (Vedensky, 1946). Alternative methods that could be evaluated are electrolysis, precipitation of manganese hydroxide with lime, and precipitation of manganese carbonate with soda ash.

Other key factors that must be researched in the field are distribution of leach solutions to thick sections of broken ore and determining the resultant recovery that can be expected. Solar evaporation ponds have not been used in commercial leaching operations and it is not known how well they could be integrated into the system. Blasting plans must be developed and their costs compared with those assumed in the economic analysis. Environmental safeguards required by regulatory agencies must be ascertained, particularly their impact on costs of an operation.

The Bureau is planning continued research to resolve some of these questions. Recent experiments on dissolving manganese oxides from sea nodules have confirmed that sulfur dioxide does leach manganese preferentially before iron and other contaminants (Khalafalla and Pahlman, 1981). Laboratory tests are being conducted on Artillery Mountain ore to better define SO_2 leaching characteristics for further evaluation of in situ leaching possibilities. Tentative plans are being made to conduct field in situ leaching experiments. Although the Bureau is not locked into an Artillery Peak site, it has to be considered a strong candidate for the same reasons that it was selected for detailed analysis. It also provides the opportunity to test certain key features of a leaching system without the complications introduced by difficult hydrometallurgy and/or complex geology.

If the research is successful, industry will be able to expand immediately into commercial exploitation of a resource that is simply too low grade to be mined any other way. And the Bureau can then use the information gained to develop leaching systems for more complex mineralogy and leaching solutions.

Summary

Although there is no commercial exploitation of manganese deposits in this Country, numerous deposits exist. The Artillery Peak deposit is an excellent example of the few large, low-grade deposits that have been known and studied for many decades. With current market conditions and mining methods, however, these low-grade deposits cannot economically compete with foreign deposits. New and innovative mining techniques are needed to achieve significant domestic production of manganese.

A relatively new technique--in situ leaching--offers possibilities for economic production from manganese oxide deposits. An economic analysis of a conceptual design for in situ leaching based upon the Artillery Peak district of Arizona showed that such a system could have a healthy rate of return and fast payoff. The Bureau is continuing research to ascertain the validity of assumptions made in the conceptual design and its evaluation.

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Table 1. Principal Manganese Deposits (After Manganese Recovery Technology Panel, 1976)

Resource	Amount of ore, million short tons	Manganese, percent	Iron, percent	Principal manganese minerals
Aroostook, Maine	377	9	20	Braunite (silicate)
Artillery Peak, Arizona	175	4	3	Wad (oxide) Psilomelane (oxide)
Batesville, Arkansas	196	4	1-3	Hausmannite (oxide) Braunite (silicate)
Chamberlain, South Dakota	>2,200	0.6	1	(Carbonates)
Cuyuna Range, Minnesota	272	8	32	Manganite (hydroxide) Psilomelane (oxide)
San Juan Mtns., Colorado	48	8	low	Rhodonite (silicate) Rhodochrosite (carbonate)

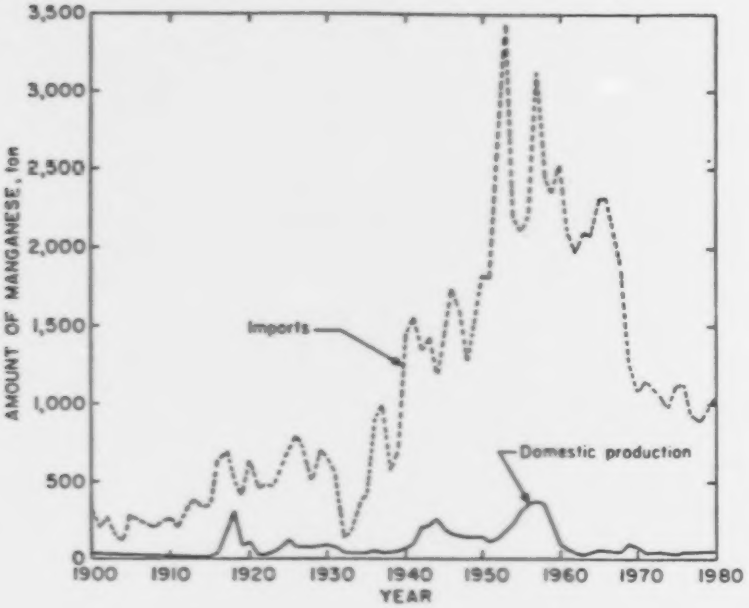

Figure 1. Manganese production trends.

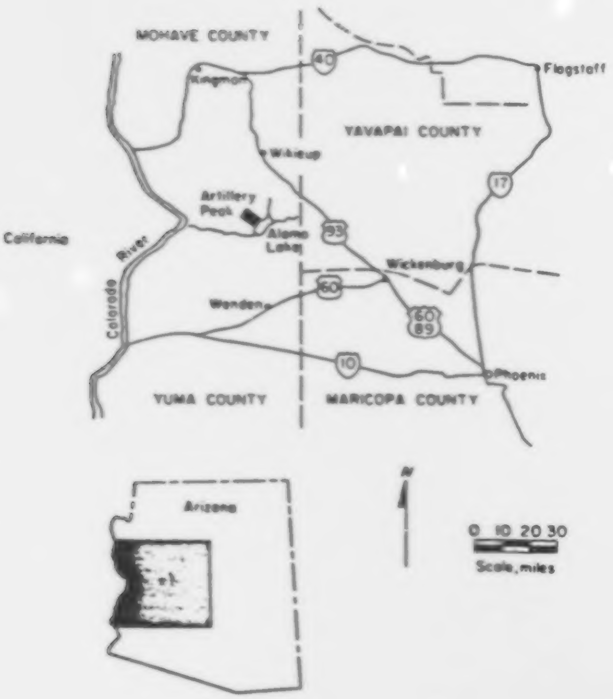
MOHAVE COUNTY
40
Kingman
Flagstaff
YAVAPAI COUNTY
Wikieup
17
Artillery Peak
Alamo Lake
93
California
Colorado River
Wickenburg
60
Wenden
60
89
10
Phoenix
YUMA COUNTY
MARICOPA COUNTY
Arizona
N
0 10 20 30
Scale, miles

Figure 2. Artillery Peak District location.

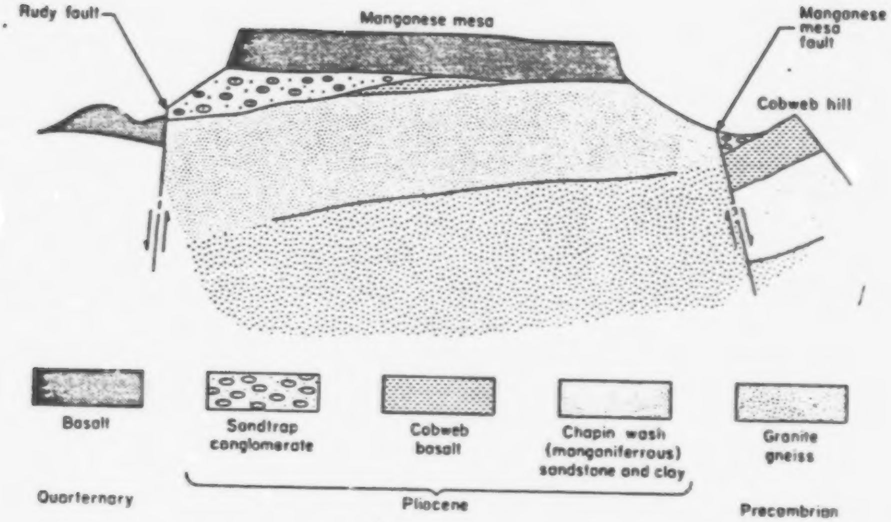
Rudy fault
Manganese mesa
Manganese mesa fault
Cobweb hill
Basalt
Sandtrap conglomerate
Cobweb basalt
Chapin wash (manganiferrous) sandstone and clay
Granite gneiss
Quarternary
Pliocene
Precambrian

Figure 3. Geologic section (east-west) of Artillery Peak District.

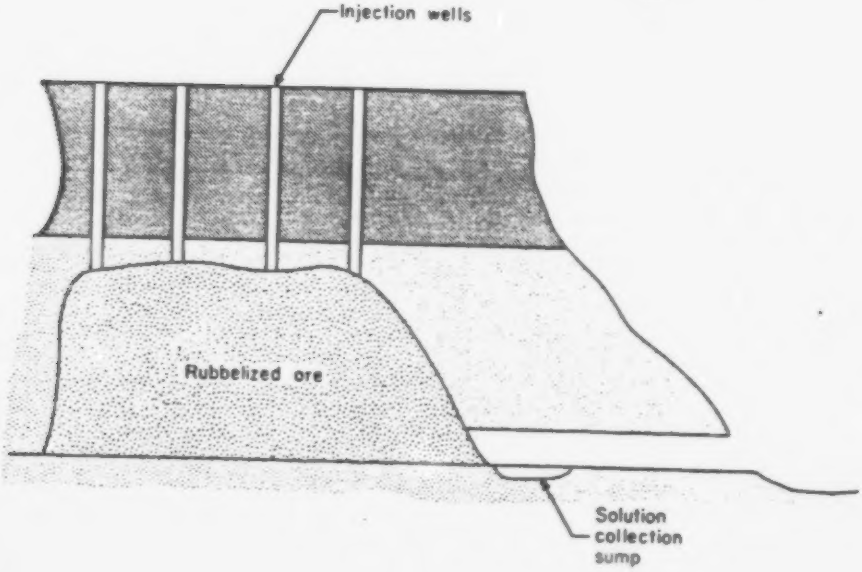
Injection wells
Rubbelized ore
Solution collection sump

Figure 5. Ore preparation for leaching.

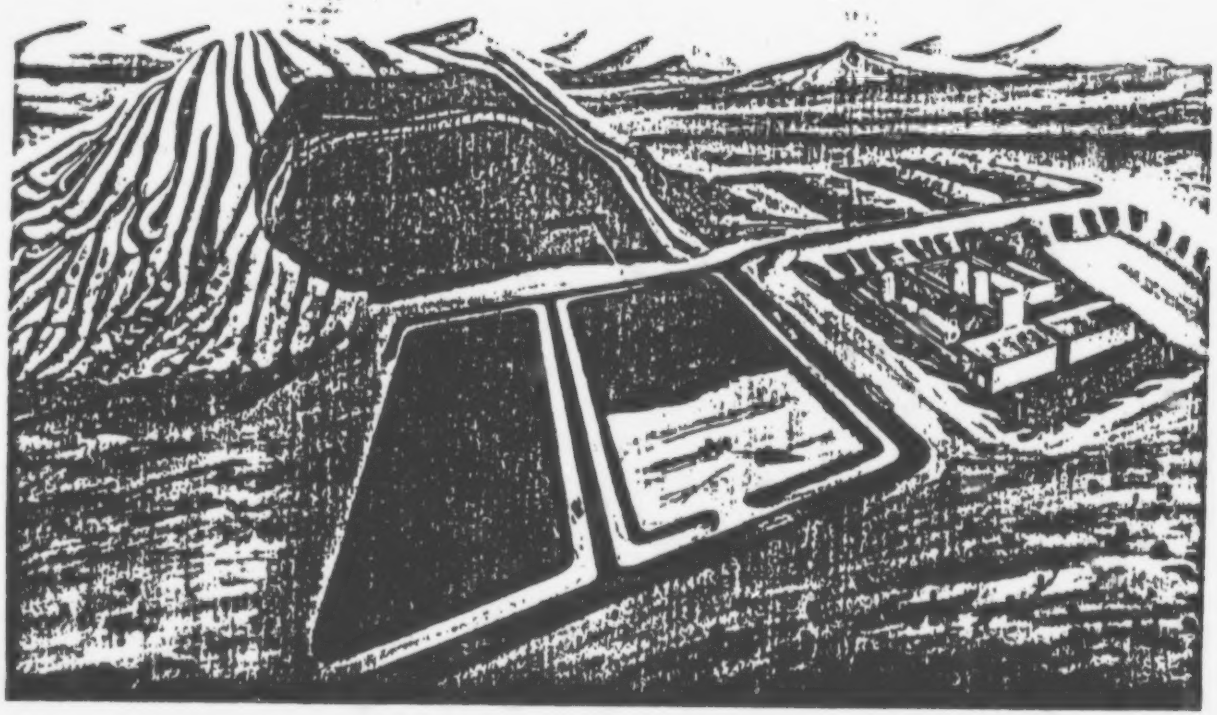

Figure 4. Conceptual design for in situ leaching.

MANGANESE

by

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When manganese additions are made in the furnace a loss of 5–10 per cent of the amount added is said to occur between tap stream and metal ingot. The addition of fine ferro-silicon to the stream in an amount required to produce the required silicon content in the finished steel substantially reduces this loss.

Whenever possible the manganese addition should be made to the ladle to minimize the loss of this element. According to Babcock, a high efficiency of manganese utilization can be obtained by a method used by German open hearth furnace operators who tapped heats into tilted ladles to reduce the mixing of air and slags with the molten steel.

Table XX gives the recovery efficiency of manganese at various stages of steel plant operation and illustrates the points described above.

Table XX. Percentage Recovery of Manganese
(*after* D. E. BABCOCK[3])

Stage	*Recovery*
Basic blast furnace	74·10
Ferro-alloy furnace	77·20
Basic open hearth	13·47
Basic ladle addition (average practice)	63·26
Basic ladle addition (special German practice)	90 +

3.2. RECOVERY OF MANGANESE FROM SLAGS AND LOW GRADE ORES

A vast quantity of manganese exists in the form of ores in which the manganese content is not sufficiently high or the manganese: iron ratio too low to favour their use for the blast furnace production of standard ferro-manganese. The large quantities of manganese contained in slag dumps associated with the steel industry have already been mentioned above. Many processes have been proposed for the production of manganese in usable form from both low grade ores and from slags. Several processes of this nature, some of which have been described in the preceding chapter, were used in Germany on quite a large scale during the war of 1939–45 to overcome the manganese deficiency there. According to Case and Clegg, several of the known techniques are almost economically competitive with the normal smelting of high grade ores to ferro-manganese. Pilot plant studies of some of these processes have already been made and large scale plants could be built and operated in a very short time in the U.S.A. if importation of high grade ores was restricted or the price increased. The processes which have been proposed may be divided into the two main classes of chemical methods and pyrometallurgical methods.

3.2.1. *Chemical processes*

Chemical processes for the treatment of low grade manganese ores have been considered at many times since the first discovery of the value of manganese to the steel industry. In a review of the patent literature given by R. S. DEAN, E. S. LEAVER and T. L. JOSEPH[4] several patents are referred to which were taken out between 1862 and 1900 although there is no record that they were commercially developed. The chemical separation of manganese from iron, silicon and other elements present in the ores is relatively simply accomplished but the development of a successful commercial process requires that such a process should use cheap materials and have saleable by-products. The processes which have been seriously considered have been those in which the ore is treated with sulphur dioxide, sulphurous or sulphuric acids, with ammonia or with nitric acid.

Sulphur processes—Manganese dioxide may be readily converted to manganese sulphate by treatment by gaseous sulphur dioxide or by sulphurous or sulphuric acid solutions at concentrations which do not convert iron oxide to sulphate. This has been the basis for several proposed processes which can be applied to ores containing manganese dioxide. Their application to carbonate ores requires the preliminary roasting of the ores to convert the carbonate to oxide and they are not applicable to ores in which manganese is present as the silicate.

Leaver drum leaching process and its later developments—In this process, first described by E. S. LEAVER[5], manganese in the ore was dissolved in hot sulphurous acid which was generated by passing sulphur dioxide gas over and through a slurry of the ore. A flow sheet for the process is shown in *Figure 24*. The ore was first wet crushed in a ball mill to 20-mesh and the wet slurry which had a consistency of about 2 parts of water to one of crushed ore was fed into a specially designed drum through which hot furnace gas containing 2–6 per cent sulphur dioxide by volume passed in the direction opposite to that of the slurry. The drum was so designed and rotated that there was intimate contact between the mineral particles and the sulphur dioxide of the gas. The ore pulp got gradually hotter as it

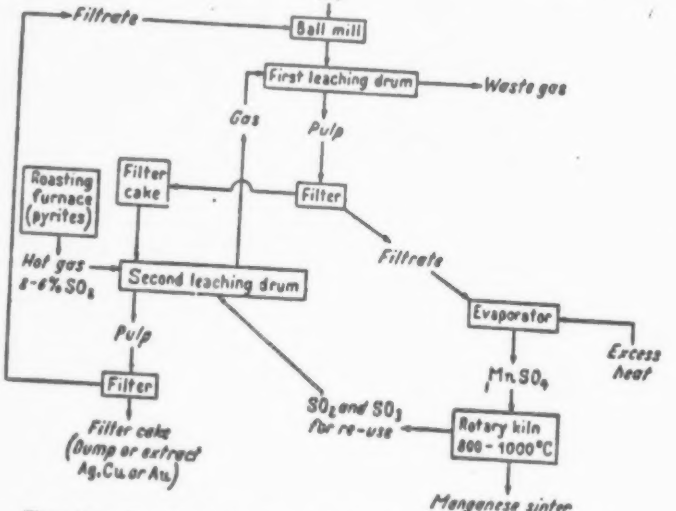

Figure 24. Flow sheet for Leaver drum leaching process (after E. S. Leaver[3]*)*

passed along the drum and left at a temperature of about 45° C, at which temperature it contained no free sulphur dioxide gas. The discharged pulp was filtered to give a solution almost saturated with manganous sulphate (nearly 11 per cent). This corresponded to a removal of about 60–70 per cent of the original soluble manganese content of the ore. The filter cake from this drum was re-pulped with fresh or wash water solution and treated in a second drum. Hot gas containing sulphur dioxide passed directly through this drum and thence to the first drum. The solution from this second drum contained 3–6 per cent manganese and this was used in the ball mill to regrind fresh ore. Some iron and phosphorus were removed along with the manganese in the second drum but these were precipitated out when the solution encountered the large quantity of ore in ball milling with a consequent neutralization of its free acid content.

The manganese was recovered from the filtrate from the first drum by evaporation and the sulphate was calcined to give an oxide product containing 60–64 per cent manganese in a rotary type furnace at 800–1000° C. The sulphur dioxide and sulphur trioxide in the gas resulting from the decomposition were re-used in the leaching drum.

A process for the treatment of manganese ores at a higher temperature than that used in the Leaver process and its later developments has been investigated on a laboratory scale by R. Blumberg and T. D. Morgan[6]. This depends upon the treatment of ore with pyrites burner gas containing 15 per cent sulphur dioxide. The conversion to manganese sulphate is exothermic. It was found that if the process was carried out at a temperature of 300° C very little iron was converted to sulphate. 60–70 per cent of the manganese present in the ore was converted to sulphate in a period of 30 minutes with almost complete utilization of sulphur dioxide in the furnace gas. Leach filtrates from the treated ore contained less than 0·02 g/l of iron and no sulphite or dithionate ions. The possibility exists of retreating the leached treated ore in order to obtain more complete extraction of manganese.

Variants of sulphur dioxide processes have been investigated on quite a considerable scale in the U.S.A. The Manganese Ore Co. in 1943 opened a plant at Las Vegas in Nevada with the intention of treating 1000 tons a day of 20 per cent manganese ore from the adjacent Three Kids mine. D. N. Vedensky[7] has described the mode of operation of the plant and the difficulties encountered in the large scale development of the process, which had previously been operated only on the scale of 3–5 tons a day.

Difficulties were experienced during leaching due to side reactions involving the formation of polythionates, notably the dithionate which was formed by the reaction

$$2H_2SO_3 + MnO_2 = MnS_2O_6 + 2H_2O$$

Thiosulphate was also formed by reaction between elemental sulphur and sulphurous acid

$$S + H_2SO_3 = S_2O_3^{-} + 2H^{+}$$

The extent to which these reactions took place was dependent upon the pH of the acid and they could be minimized by keeping the pH below about 3·5.

Another major difficulty arose from the formation during leaching of calcium sulphate which was particularly objectionable because of its scale-forming tendency. It caused clogging of baffles in the leaching towers and interfered with gas scrubbing consequently leading to excessive loss of sulphur dioxide, unless some sacrifice of manganese extraction efficiency was made.

Further difficulties were encountered during evaporation and in separation of the manganese sulphate and dithionate which it was necessary to achieve unless a serious loss of sulphur was to be accepted.

Finally trouble was caused in the roasting kilns. Large amounts of alkali salts went to the roaster because of the above difficulty in separating manganese and alkali salts in the separators. These alkali salts formed a low melting point eutectic with manganese sulphate but the melting point subsequently rose as the sulphur dioxide was driven off. This resulted in the formation of hard rings in the kiln which were difficult to dislodge, and caused the kilns to be shut down frequently.

Although many of these difficulties were overcome or at least controlled during the operation of the plant, the production rate did not exceed 50 per cent of that projected and the plant was shut down after operating for about a year due to high operating costs and the development of a favourable supply position for foreign ore.

A recent process which has been judged to have some promise has been developed by the Chemical Construction Corporation and is called the Chemico Process. It has been described by L. N. Allen[8]. It differs from early sulphur dioxide processes in two respects—a high temperature autoclaving step and a sintering operation. As in other processes, the ore concentrates are treated in slurry form with sulphur dioxide gas to convert manganic oxide to manganese sulphate but the reacted slurry is then heated in an autoclave in the presence of oxygen. This is claimed to improve the extraction efficiency and to eliminate by oxidation the dithionate which was so troublesome in the Three Kids plant. The manganese sulphate solution is decanted from the autoclave and the tailings are washed to recover soluble salts. The solution is then evaporated, the manganese sulphate crystals separated and sintered to manganese oxide, the sulphur dioxide gas produced during sintering being recycled. The process is said to be suitable for the treatment of low grade ores or concentrates containing about 20 per cent manganese and to produce a sinter containing 60 per cent manganese and with a low sulphur content.

Detailed cost estimates for large scale production by the Chemico Process have been made and independently checked but the process has not yet been put into commercial operation. According to Case and Clegg, the estimated costs come very near to making the product competitive with high grade ore imported into the U.S.A. Tests made by the Battelle Memorial Institute on the process indicate recoveries of the order of 90–95 per cent of the manganese and it has been estimated that the cost of high oxide sinter produced by this process should be about $100 per long ton of contained manganese.

The dithionate process—As an alternative to the later stages of the processes described above an attempt was made in early investigations at the U.S. Bureau of Mines to avoid the calcination stage by treatment of the manganese sulphate solution with calcium chloride solution to precipitate calcium sulphate and form manganese chloride solution from which the manganese was precipitated as a hydrated oxide by treatment with calcium hydroxide, and the calcium chloride solution regenerated.

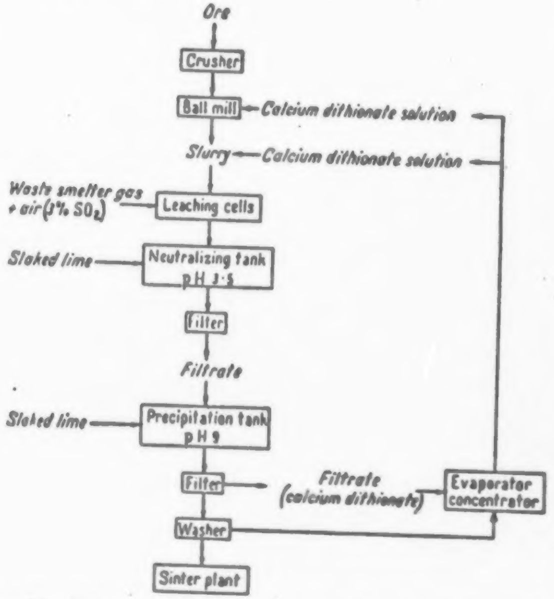

Figure 25. Flow sheet for dithionate leaching process (after S. F. Ravitz et al.[9])

During the investigation of this process it was found that as the liquors were recycled oxidation of sulphur dioxide by manganic oxide to dithionate ion occurred so that the calcium chloride content of the solution gradually decreased while the calcium dithionate content increased. Since the latter was less corrosive than calcium chloride solution and obviated the necessity for the latter a process was developed based entirely on calcium dithionate. The development of this process has been described by S. F. Ravitz, W. F. Wyman, A. E. Back and K. E. Tame[9]. A flow sheet for the process is given in *Figure 25.*

The basis of the process is that a slurry of the ore in an excess of dithionate solution is treated with a gas, such as waste smelter gas containing 3 per cent sulphur dioxide and oxygen, leading to the following reactions:

$$MnO_2 + SO_2 = Mn^{2+} + SO_4{}^{2-} \quad .... (1)$$

$$MnO_2 + 2SO_2 = Mn^{2+} + S_2O_6{}^{2-} \quad .... (2)$$

$$H_2O + SO_2 + \tfrac{1}{2}O_2 = 2H^+ + SO_4{}^{2-} \quad .... (3)$$

$$H_2O + 2SO_2 + \tfrac{1}{2}O_2 = 2H^+ + S_2O_6{}^{2-} \quad .... (4)$$

The sulphate formed by reactions (1) and (3) is precipitated as calcium sulphate by reaction with calcium dithionate by the reaction

$$Ca^{2+}S_2O_6{}^{2-} + SO_4{}^{2-} = CaSO_4 + S_2O_6{}^{2-} \quad .... (5)$$

After filtration to remove gangue and precipitated calcium sulphate the solution is treated with slaked lime to precipitate manganese hydroxide and regenerate calcium dithionate

$$Mn^{2+}S_2O_6{}^{2-} + Ca(OH_2)_2 = Mn(OH)_2 + Ca^{2+}S_2O_6{}^{2-} \quad .... (6)$$

The precipitate is filtered off and sintered to produce manganese dioxide and the calcium dithionate is recycled. Ravitz *et al.* point out that the only losses of dithionate are those due to incomplete washing of the leach residue and the precipitated hydroxide. To maintain the dithionate concentration in the circuit, the quantity of dithionate formed by the above reactions (2) and (4) should equal these losses.

This process was operated on a semi-pilot plant scale treating 400 lb of ore daily. The ores contained 13–18 per cent manganese and using an initial concentration of dithionate ion in the feed solution of 180–200 g/l extractions of manganese from the ores ranging from 85–95 per cent were attained. 2–4 per cent of the extracted manganese was lost by co-precipitation with calcium sulphate and 0·6–1·5 per cent was lost as soluble manganese in the washed residue. The recovery was thus 96 per cent of the manganese extracted.

The total amount of lime required ranged from 300–430 lb per ton of ore leached depending upon the type of ore. The utilization of sulphur dioxide was 60–70 per cent of that introduced and ranged from 1·5–2·2 lb per lb of manganese extracted. Since this came from smelter gas which would otherwise be wasted its contribution to the extraction cost was negligible but, except under such conditions, the conversion of sulphur dioxide to a relatively worthless by-product, calcium sulphate, would be a major drawback to this process.

The ores contained up to 3·5 per cent iron and 0·1–0·2 per cent phosphorus. After sintering the percentage analysis of the product was: Mn 61, Fe 1·0, P 0·04, SiO_2 0·7, CaO 7·3. Sulphur present to the extent of 2–3·5 per cent in the precipitated oxide, mainly as sulphur trioxide, was reduced by sintering to 0·06 per cent. Such material would meet all the specifications for ferro-grade ores.

Ferrous sulphate process—An essential operation in the steel industry is pickling in dilute sulphuric acid to remove mill scale. The waste pickle solutions containing 0·5–10 per cent sulphuric acid and 10–22 per cent ferrous sulphate by weight are discarded. It has been estimated that 600,000,000 gallons of this waste solution are produced annually in the U.S.A. alone. The disposal of this solution is often a major problem. R. D. HOAK and J. COULL[10] have proposed a process for the use of waste pickle solution to leach manganese from low grade ores. A flow sheet for the process is shown in *Figure 26*. Experimental work with low grade ores containing 14·7–26·9 per cent manganese showed that, when ground to

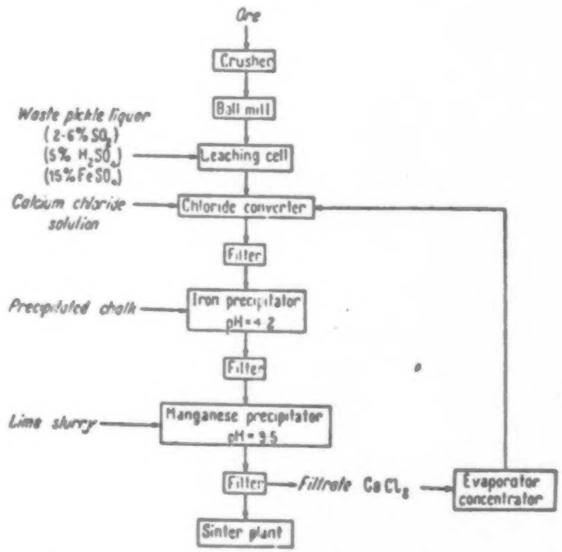

Figure 26. Flow sheet for ferrous sulphate leaching process (after R. D. HOAK *and* J. COULL[10])

about 60-mesh, as much as 98 per cent of the manganese could be extracted in a pickle liquor of average composition (5 per cent free sulphuric acid and 15 per cent ferrous sulphate by weight) in 30 minutes. Filtration of the extract gave a solution containing ferrous and manganese sulphates.

As in the dithionate process described above, it is desirable to convert the sulphates to chlorides before precipitation of the manganese in order to minimize contamination of the manganese with sulphur. The next stage in the process was therefore to convert the sulphates to chlorides. This was achieved by treating the solution with calcium chloride solution. This, however, has the disadvantage of producing calcium sulphate dihydrate as a by-product of relatively low value.

Separation of the iron and manganese from the filtered chloride solutions was achieved by selective precipitation of the iron. This could be obtained by the addition of a slaked lime slurry or precipitated chalk. The latter was found to be more effective in minimizing co-precipitation of manganese with iron and it was found that if the chalk was added slowly and with violent agitation and if the average pH of the mixture was not allowed to rise above 4·2, 98 per cent of the iron could be precipitated carrying no more than 5 per cent of the manganese with it.

The filtered extract after precipitation of the iron consisted of a solution of manganous and calcium chlorides. From this the manganese was precipitated as hydrated oxide by the addition of lime slurry until the pH rose to 9·5. The precipitate was filtered off while the filtrate was concentrated by evaporation and recycled for re-use in the chloride conversion step. The precipitated oxide obtained from an ore containing 16 per cent manganese contained 60·9 per cent manganese and, by roasting, this was raised to 64·5 per cent. The silica content of the unroasted oxide was 1·84 per cent and the iron content 2·33 per cent. The sulphur content, as SO_3, was 1·86 per cent but, although data were not given by Hoak and Coull, it may reasonably be inferred that this could be reduced by roasting to a value as low as that reported for the concentrate produced by the dithionate process described above, *viz* less than 0·1 per cent.

Although the process as described has the attractive feature of utilizing an otherwise wasted commodity in the shape of waste pickle liquor, a considerable disadvantage is the fact that in general in most countries the steel works which are the source of the pickle liquor are at a considerable distance from the deposits of low grade ore. Expensive transportation is therefore involved in bringing one commodity to the other and this may outweigh the advantages of the process and make it economically unprofitable.

Ammonia processes—(*i*) *Bradley–Fitch process.* Like the process described in the preceding chapter for obtaining an electrolyte for production of electrolytic manganese from manganese ore, the ammonia process requires a preliminary reducing roast to convert the manganese dioxide MnO_2, which is insoluble in ammonia, to manganous oxide MnO, which is soluble, while iron oxide FeO, which is also soluble, is converted during roasting to the magnetic oxide Fe_3O_4, which is insoluble. The Bradley–Fitch process was operated on a pilot plant scale with a capacity of about 4 tons a day at the Minnesota School of Mines. The operation of this plant has been described by Dean, Leaver and Joseph[4]. A simplified flow sheet for the process is given as *Figure 27*. The ore which was treated had the percentage composition: Mn 17, Fe 29·93, SiO_2 24·42, Al_2O_3 4·14, CaO 1·0, and P 0·06. This was crushed and ground wet to

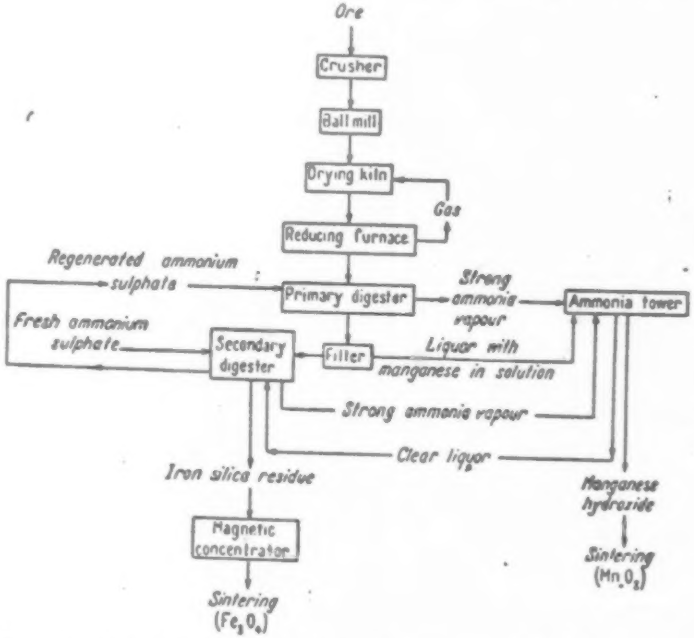

Figure 27. Simplified flow sheet for Bradley–Fitch ammonia process [*after* R. S. DEAN *et al.*; *reprinted from* Inform. Circ. U.S. Bur. Min. No. 6770 (1934)]

standard ferro-manganese from high grade ore was $185 per gross ton. The process thus appears to have considerable economic promise.

Obviously by processes such as these a major industrial country such as the U.S.A. having access to ferruginous manganese ores or manganiferous iron ores could make itself self-sufficient in manganese if cut off from overseas supplies of high grade ore. Under normal conditions, however, the development of any of the processes described in this chapter will be gradual and will depend upon accurate cost evaluations based upon further pilot plant studies. Obviously the position is very delicately balanced since several of these processes show promise of economic operation and could easily become decisively so if there was any world wide increase in the cost of high grade ore as a consequence of increased mining or transportation costs.

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